CONCEPTUAL DESIGN AND TECHNO-ECONOMIC ASSESSMENT OF THREE ELECTRICITY-BASED REACTOR PROCESSES FOR PRODUCTION OF LIQUID FUELS AND/OR HYDROCARBONS FROM NATURAL GAS

by
Technical Development Department
SNC Inc.

for
CANMET
Energy, Mines and Resources
Canada

March 1985
Montréal

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DSS Contract No.	:	OSQ83-00070
DSS File No.	:	265Q.23440-3-9076
CANMET Project No.	:	423351
CANMET Scientific Authority	:	D.P.C. Fung
SNC Project No.	:	7073

The study was funded 50/50 by CANMET and SNC Inc.

128 pp

AUG 19 1988
MICROMEDIA

March 1985
Montréal

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ACKNOWLEDGEMENTS

It will be evident from the content of the report that substantial information and data was provided by several process developers from USA, Federal Republic of Germany, Norway and Canada. In random order, assistance provided by Dr. R. Muller (Chemische Werke Huels); Dr. S.L. Thurnberg and Mr. W.H. Reed (Westinghouse Electric Corporation); Dr. E. Sodahl and Dr. A. Holmen (SINTEF); Messrs. G.J. Weiss and W.L. Nolan (Mobil Research and Development Corporation); Mr. B. Mori (Union Carbide Canada); D.L. Vanstone (Trans Alta Utilities); and Dr. J. Happel (Catalysis Research Institute) is gratefully acknowledged.

Appreciation is also expressed to Dr. D.P.C. Fung, the Scientific Authority appointed by CANMET for the study, for continued cooperation, and input to the study.

snc

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EXECUTIVE SUMMARY

This cost-shared study has been carried out by the Technical Development Department of SNC Inc. for the Department of Energy, Mines and Resources, CANMET, to investigate the feasibility of liquid fuel (gasoline) and/or hydrocarbon (acetylene and ethylene) production from natural gas via the novel acetylene route using the following designated electricity-based acetylene reactors:

- electric-arc reactor (Chemische Werke Huels, Germany)
- preheated or hydrogen plasma reactor (AVCO, Westinghouse, USA)
- resistance-heated type (Happel and Kramer, USA; and SINTEF, Norway).

Canada has surplus natural gas and in some provinces, particularly Quebec, cheap electricity is also available. The proposed scheme involves utilization of both natural gas and electricity to produce the desired products.

Apart from mechanical divergencies, the designated acetylene reactors also differ in the way the electrical energy is transferred to the feedstock. However, a common principle characterizes the designated reactor systems considered in the study. Typically, natural gas is heated rapidly to a high temperature (1500-1800 C) in less than 5 milliseconds at an operating pressure between 40-60 kPa. The product gas leaving the reaction zone is rapidly quenched (< 2ms) with LPG to stabilize the acetylene formed and to suppress the undesirable carbon formation.

The **"electric-arc"** reactor of Chemische Werke Huels transfers electrical energy by "direct" contact between the high-temperature arc (15 000-20 000 C) and the hydrocarbon feedstock. Stability of the arc is achieved by rotating the feed gas through a swirl chamber at the reactor inlet. The process has been commercialized for many years. Operational difficulty experienced earlier due to deposition of hard graphite on the electrodes have been successfully overcome.

In the **"preheated hydrogen"** (commonly known as **"hydrogen-plasma")** reactor, electrical energy is transferred "indirectly" to the feedstock. Hydrogen is fed into the arc zone and heated to a plasma state. The existing stream of hot hydrogen plasma at about 5000 C is mixed rapidly with the incoming feed below the arc zone, and undesirable carbon formation on the electrodes is therefore minimized. This concept has been tested by Chemische Werke Huels for heavier hydrocarbon feedstock, by AVCO for coal and by Westinghouse for natural gas.

The **resistance-heated** reactor proposed by Happel and Kramer and further developed by SINTEF consists of concentric tubes (carbon resistors) heated by passage of current. The narrow annular space is the reaction zone. The temperature of the reactor wall is kept between 1800-1900 C to provide the temperature of 1600-1800 C in the annular reaction zone. The "isothermal" wall reactor provides better control of temperature profile, thus allowing conditions which provide higher yields of acetylene and lower carbon formation.

For each of the reactor configurations, a conceptual design with overall material, energy and utility balances has been prepared for a commercial plant producing 12 500 bbl/d of liquid fuel (gasoline) product. The overall plant consists of an acetylene reactor system, acetylene hydrogenation, olefin oligomerization (Mobil Olefin to Gasoline and Distillate Process) and hydrogen recovery (Union Carbide's Pressure Swing Adsorption) units along with the required support, utility and offsite facilities to comprise an integrated, self-sufficient operation. Natural gas, electricity, LPG (quenchant) and make-up water are the major process inputs. Gasoline is the main product. Hydrogen is recovered as a by-product.

Prescreening or budget-type capital, operating and product cost estimates were prepared for the 12 500 bbl/d gasoline plant using each of the three designated acetylene reactor configurations. Equipment cost and investment data supplied by various process developers from Germany, USA and Norway were adjusted and converted for Canadian conditions using approved techno-economic criteria and consistent procedures developed for the study. Where information was not available, factored-cost estimates or SNC in-house data were used. Some of the basic techno-economic criteria are summarized below:

- Plant location : Quebec
- Constant Canadian dollars: mid-1984 basis
- Construction period : 3 years
- Financing : 100% equity
- Cost of capital : 5% after taxes or about 9% before taxes
- Plant life : 20 years

- Depreciation : Straight line with zero salvage value
- Plant service factor : 8000 hours/yr (0.91)

Cost of natural gas was taken at \$3.86/GJ (\$4.0/10^6Btu), LPG at \$166/m^3 (16.6¢/litre), and electricity cost at 1¢/kWh in view of large consumption. Byproduct credits for hydrogen were taken at \$4.74/GJ (\$5.0/10^6 Btu) and \$7.58/GJ (\$8.0/10^6 Btu).

A summary of techno-economic results is presented in Table 1. Conversion of natural gas is in the range of 80-84%, being highest for the resistance-heated reactor. Similar trends are observed for methane conversion, acetylene product yield and the specific energy requirements (kWh/kg of acetylene formed). However, the overall process SER is nearly the same for the resistance-type and electric-arc reactor ($\simeq$ 10 kWh/kg) of gasoline); though for the hydrogen plasma reactor, the SER is somewhat higher (12.5 kWh/kg of gasoline).

Similarly, the overall thermal efficiency of electric-arc and resistance-type configurations is nearly the same (about 61%); though for the hydrogen-plasma reactor it is somewhat lower (54%). The liquid product effeciency is, however, nearly the same for all three reactor configurations (42-43.6%).

The resistance-type reactor has been tested only on bench-scale and pilot plant levels; therefore, significant research and development may be required prior to commercialization. At the present level of development, a high degree of technical risk is involved. Also, undesirable carbon formation in the narrow annular space may lead to significant operational problems. The

Table 1-0 Techno-Economic Comparison of the Designated Processes
Plant Capacity: 12 500 bbl/d gasoline product

	Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Natural Gas Conversion, wt%	79.5	79.5	83.0
CH_4 Conversion[1], Prequench, wt%	82.0	82.0	88.0
After quench*, wt%	74.0	74.0	80.0
Reactor Product Yield (Pre LPG Quench), wt%			
. Acetylene Yield[1]	58.5	66.6	79.7
. Acetylene + Ethylene Yield[1]	68.0	77.5	83.7
Reactor SER (Pre LPG Quench)			
. kWh/kg of acetylene	11.3	14.5	9.9
. kWh/kg of acetylene + ethylene	9.7	12.5	9.4
Reactor SER (After LPG Quench)**			
. kWh/kg of acetylene	11.3	14.5	9.9
. kWh/kg of acetylene + ethylene	8.0	10.3	8.0
Process Input, kg/kg of gasoline			
. Natural Gas	1.30	1.19	1.17
. LPG	0.23	0.20	0.25
Overall Process SER, kWh/kg of gasoline	10.0	12.5	10.2
By-Products Output, kg/kg of gasoline			
. Hydrogen	0.17	0.13	0.14
Total Energy Input, GJ/h	6 909	7 040	6 594
Total Energy Output, GJ/h	4 169	3 819	4 067
Thermal Efficiency, %			
. Overall	60.3	54.2	61.7
. Liquid	41.8	41.0	43.6

* Lower overall CH_4 conversion (after quench) is due to some methane formation during LPG quenching.

** Reactor SER (after LPG quench) based on acetylene yield is the same as SER (pre LPG quench) since negligible acetylene is formed during quenching.

Table 1-0 Techno-Economic Comparison of the Designated Processes
Plant Capacity: 12 500 bbl/d gasoline product (Cont'd)

	Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Development Status	Commercial	Pilot Plant	Bench-Scale Pilot Plant
Capital Investment, $ million *			
. Direct	522.91	539.00	487.09
. Indirect	35.19	36.28	33.12
Total	558.10	575.28	520.21
Investment, $/daily litre of gasoline	280.8	289.5	261.8
Gasoline Product Cost, ¢/litre			
. At $7.58/GJ H_2	37.5	41.3	36.0
. At $4.74/GJ H_2	41.9	44.5	40.1
Breakdown of Product Cost, %			
. Process Input[2]	84.0	75.3	83.6
. Operating Cost[3]	15.2	13.2	15.0
. Overheads[4]	7.2	6.8	7.2
	106.4	95.6	105.8
. By-products Credits	(31.4)	(20.6)	(30.3)
	75.0	75.3	75.5
. Depreciation	10.9	10.2	10.6
. Capital Recovery Cost	14.1	14.5	13.9
	100.0	100.0	100.0

* Includes 15% of installed plant cost for engineering.

Notes: 1. Based on total CH_4 feed to the reactor (i.e. including CH_4 in the recycled gas stream).

2. Process input includes natural gas, LPG and electricity.

3. Operating cost includes catalyst and chemicals, make-up water, running royalties, labor, supervision, operating supplies and maintenance cost.

4. Overheads include property taxes, insurance, general overhead and administration expenses.

electric-arc reactor is commercially proven. The hydrogen plasma reactor has also been extensively tested,but further development and demonstration on a PDU and pilot-plant scale is required. Cost of development of hydrogen plasma and resistance-type reactors are not added to the estimated capital cost. These costs are likely to be in the range of $10-15 million dollars, which will have a negligible impact on the final product cost.

The novel scheme for liquid fuel production via the acetylene route offers an attractive synfuel approach for liquid fuel and/or hydrocarbon production. The proposed route is less capital intensive. The capital investment of $260-280 per daily litre of gasoline product is less than 50% compared to other synfuel technologies such as Fischer-Tropsch or conventional coal liquefaction processes.

The product cost (transfer price) of gasoline ranges between 36-41.3¢/litre depending on the reactor configuration at $7.58/GJ hydrogen byproduct credit. The major cost component is raw material and process input (natural gas, LPG and electricity) accounting for about 80% of the net product cost. Other significant components are byproduct credits for hydrogen and capital recovery cost.

Although the scope of the study did not include the sensitivity analysis, a possible 25% reduction in the cost of natural gas from the full border price of $3.86/GJ would reduce the gasoline product cost by 5¢/litre. If, however, a 20% higher electricity cost (i.e. 1.2 ¢/kWh) is considered, the final gasoline product cost would increase by 1.1 to 1.8¢/litre depending on the reactor configuration. Other industry incentives, e.g. tax credits,

which would have further positive impact on the final product cost were not considered at this stage of the study. Compared to existing cost of gasoline (32-35¢/litre) using conventional technology, the estimated cost of gasoline from the proposed route is attractive. It is recommended that further study be done to explore the potential of making gasoline via the acetylene route.

As a sub-objective of the study, preliminary capital, operating and product costs were calculated for designated hydrocarbon production (acetylene and ethylene). Two plant capacities were considered for acetylene production: a 50 million lb/yr (22.7 Gg/yr) acetylene to be used as a feedstock for specialty chemicals such as butanediol, n-methyl-2-pyrrolidone, etc., and a 300 million lb/yr (136.2 Gg/yr) acetylene plant for commodity chemicals production such as vinyl chloride, acrylonitrile etc. For ethylene production, a plant capacity of 1 billion lb/yr (454 Gg/yr) was considered.

The results are summarized in Table 2. The calculations are shown for the electric-arc reactor configuration. For the hydrocarbon option, byproduct credits were taken for hydrogen at \$4.74/GJ (\$5.00/10^6 Btu), benzene at \$540/t (US \$1.40/US gal), ethylene at 55¢/kg (25¢/lb) and propylene/butylene at 42¢/kg (19¢/lb).

Product cost (transfer price) of acetylene for a 50 million lb/yr plant is estimated at 96¢/kg. Total investment required is approximately \$81 million. At this cost, co-production of acetylene is very attractive. A possible discount of 10% on

Table 2-0 Capital, Operating and Product Cost for Hydrocarbon Production
Plant Location: Quebec
Reactor Configuration: Electric-Arc (Chemische Werke Huels)

	Millions of Canadian Dollars (mid-1984)		
	Acetylene Plant		Ethylene Plant
Plant Capacity million lb/yr	50	300	1000
Direct Capital Investment			
Installed Plant Cost (TPC)	67.80	239.76	456.20
Project/Estimation Contingency	6.78	23.98	45.62
Total Installed Plant Cost (TIPC)	74.58	263.74	501.82
Indirect Capital Investment			
Licence and Prepaid Royalties	1.37	1.37	2.70
Initial Catalysts & Chemicals	0.10	0.20	0.51
Start-Up and Other Costs	2.23	7.91	15.05
Sub-Total	78.28	273.22	520.08
Working Capital*	2.24	7.91	15.05
Land*	0.54	0.65	1.30
Total Capital Investment (TCI)	81.06	281.78	536.43
Annual Operating, Maintenance and Production Costs			
Raw Material and Process Input	16.04	92.8	214.72
Operating Cost	4.45	13.0	26.77
Overheads	2.68	9.15	17.39
	23.17	114.95	258.88
By-products Credits	(10.74)	(64.42)	(83.52)
	12.43	50.53	175.36
Depreciation	3.91	13.66	26.00
Capital Recovery Cost (ROE, taxes, etc.)	5.34	16.45	41.92
	21.68	80.94	243.28
Product Cost			
¢/lb	43.4	26.9	24.3
¢/kg	96.0	59.4	53.6

* non-depreciable items

natural gas would further reduce the acetylene product cost by 6¢/kg. Currently Acetylene is currently available at $1.25-1.60/kg.

For large-volume commodity chemicals production, the product cost of acetylene is calculated as 60¢/kg. At this price, acetylene is competitive with ethylene.

Production of ethylene via acetylene is not economically attractive. The product cost of ethylene at 55¢/kg is significantly higher compared to the cost of ethylene obtained from the conventional process (37-44¢/kg).

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1.0 INTRODUCTION

1.0 INTRODUCTION

This study has been carried out by the Technical Development Department of SNC Inc. as a cost-shared contract (No. OSQ83-00070, DSS File No. 26SQ.23440-3-9076) for the Department of Energy, Mines and Resources, CANMET, Canada entitled "Conceptual Design and Techno-Economic Assessment of Three Electricity-Based Reactor Processes for Production of Liquid Fuels and/or Hydrocarbons from Natural Gas".

The main objectives and scope of the study are:

- to investigate the feasibility of liquid fuel (gasoline) production from natural gas via the acetylene route using the following designated electricity-based reactor configurations:
  - "electric-arc" reactor of Chemische Werke Huels, Germany.
  - "preheated or hydrogen plasma" reactor as used by AVCO and Westinghouse, USA.
  - "resistance-heated" reactor developed by Happel and Kramer, USA and SINTEF, Norway.
- to develop a conceptual design of a liquid fuel plant producing 12 500 bbl/d (1 987 350 litres/d) of gasoline using each of the three designated acetylene-reactor configurations
- to calculate operating, maintenance and product costs for gasoline production via the acetylene route
- to carry out a preliminary techno-economic assessment and comparison of the designated processes

1.0 INTRODUCTION (Cont'd)

As a sub-objective of the study, preliminary capital, operating and product cost estimates were also developed for the following cases using the electric-arc acetylene reactor of Chemische Werke Huels for the production of:

- acetylene as a feedstock for:
  - Reppe-type chemicals, 50 million lb/yr (22.7 Gg/yr) acetylene
  - Commodity chemicals, 300 million lb/yr (136.2 Gg/yr) acetylene
- ethylene as feedstock for petrochemical production, 1 billion lb/yr (454 Gg/yr) ethylene.

Canada has surplus natural gas and in some provinces, particularly Québec, abundant as well as cheap electricity is also available. Present estimates of the established reserves of marketable natural gas from all non-frontier areas are 76 EJ or about 72 trillion cubic feet (Edge 1981). In addition, substantial gas reserves are known to exist in the frontier as well as the offshore areas near Sable Island.

Various energy options have been considered and one of the strategies has been to conserve energy and encourage utilization of natural gas for oil. The use of electricity-based methods for conversion of natural gas to liquid fuels and/or hydrocarbon products via acetylene has two main advantages:

- combined use of both surplus natural gas and cheap electricity to produce valuable products

- decrease Canada's dependence on imported oil, particularly for Eastern Canada, by shifting utilization of huge shut-in gas reserves available in the country.

The concept of converting natural gas or any hydrocarbon feedstock to intermediate acetylene using electricity-based methods is not new. In fact, the first electric-arc acetylene process was pioneered by BASF in 1920 and a commercial plant was in operation at Huels, Germany in 1940. In the United States, acetylene manufacture by this method was initiated in 1940s by Monsanto Chemical Co. for their plant in Texas City producing about 110 million lb/yr (50 Gg/yr) of acetylene. In the following years, expansion was rapid and several acetylene plants were built all over the world until about mid 1960 when acetylene was gradually replaced by cheaper ethylene as a feedstock for petrochemical production. The worldwide annual production of acetylene from hydrocarbons was estimated at 1 100 000 t/yr (2.4 billion lb/yr) in 1965 and 800 000 t/yr (1.8 billion lb/yr) in 1976 (Miller 1966).

An excellent review of different acetylene processes is made by Miller (1966) and Tedeschi (1982). In general, all hydrocarbon acetylene processes differ essentially in the manner in which necessary energy for the reaction is supplied. An advantage of the electricity-based method for acetylene manufacture is that energy could be applied very rapidly to the reactant and the conversion of hydrocarbon to acetylene is significantly higher than other conventional thermal processes, namely the Wulff's

regenerative, one - or two-step partial oxidation processes (Tedeschi 1982). The residence time is relatively short, in the order of milliseconds. Thermodynamics and kinetics considerations favoring the use of electricity-based methods are summarized in Appendix A.

The disadvantage of electricity-based acetylene processes is that energy requirement is high, consequently the method is only feasible where electricity is cheap and available. Also, these processes are sensitive to process variations, which could lead to significant amounts of undesirable carbon if poorly operated. E.I. Dupont de Nemours and Company (Doukas 1963, Cichelli 1963) operated a modified "rotating" arc process at Montague, Michigan to provide acetylene for chloroprene (neoprene rubber) manufacture. The Dupont facility, designed to produce cheaper acetylene than the Calcium Carbide process, was closed down by 1973 due to poor economics and operating problems.

Of the three acetylene reactors designated for the study, the electric-arc process of Chemische Werke Huels transfers electrical energy by "direct" contact between the high-temperature arc and the hydrocarbon feedstock. The process has been in commercial operation for over forty five years. Improvements in operating conditions have been realized and other operating problems resolved. Disturbances in the arc, which were experienced earlier due to deposits of hard graphite on the electrodes, have been successfully overcome.

At present, Chemische Werke is operating 19 reactors in a peak-shaving mode to produce about 265 million lb/yr (120 000 t/yr) acetylene, 110 million lb/yr (50 000 t/y) ethylene and

500 million Nm^3/yr hydrogen as a byproduct. The acetylene is used for vinyl chloride monomer, butanediol and trichloroethylene production (Muller 1983).

In the **"preheated hydrogen"** (commonly known as **"hydrogen-plasma")** acetylene reactor, electrical energy is essentially transferred "indirectly" to the hydrocarbon feedstock. Hydrogen is fed into the arc zone and heated to a plasma state. The existing steam of hot hydrogen plasma is used to provide heat to the hydrocarbon feedstock; this way undesirable carbon formation on the electrodes is minimized.

Other inert gases such as argon and nitrogen have been used (Leutner 1961, Miller 1966) but hydrogen is particularly attractive because it also acts as a suppressant for acetylene decomposition. Also, hydrogen is formed as a by-product with the acetylene, therefore use of hydrogen gas does not add any extra processing steps during the product separation and recovery operation. The reaction $H_2 \rightarrow 2H$ is highly endothermic (ΔH = 429 kJ/mol*), therefore, on cooling from 5000 C to acetylene reaction temperature, hydrogen plasma provides sufficient heat to form acetylene from methane. Other advantages of the hydrogen plasma reactor are described in Section 3.3. The drawbacks are that the hydrogen plasma reactor requires relatively more electrical energy input than the electric-arc process and somewhat higher investment due to higher volume of gas to be processed downstream.

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* mol = refers to gram moles

Two other methods of transferring electrical energy to the hydrocarbon feedstocks have been practised. The first is the heating of solid or liquid resistors by passage of current, the heat being transferred by contact, conduction and radiation. Secondly, heat may be provided by generation of electromagnetic radiation of appropriate frequency or induction heating of a solid in contact with the gas. The latter method is not considered in this study.

The **resistance-heated** acetylene reactor proposed by Happel and Kramer (1964) and developed further by Scientific and Industrial Research Foundation, SINTEF (Helgesen et al 1967) falls in the former category. The initial reactor consisted of a thin aluminum oxide tube heated by a surrounding graphite heating element. Other improvements have since been incorporated. The temperature of the reactor wall is kept between 1800-1900 C with the result that the temperature in the reaction zone is maintained at approximately 1600-1800 C. The electrically heated resistance tube provides better control of the temperature profile in the reactor, thus creating conditions which provide higher yields of acetylene and lower carbon formation. This reactor has been tested on bench-scale and PDU levels only.

In addition to the objectives highlighted earlier, the study was also aimed at identifying whether significant economic incentives exist for a particular type of reactor configuration over another, coupled with existing commercial processes to produce gasoline from natural gas via acetylene. An overview of available commercial or developing processes for conversion of acetylene and ethylene to liquid fuels is described elsewhere (Cooke et al 1984).

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR TECHNOLOGY ASSESSMENT AND PROCESS SELECTION

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR TECHNOLOGY ASSESSMENT AND PROCESS SELECTION

Technical and economic criteria for technology assessment and selection of processes for production of transportation fuel from natural gas via acetylene are given below. Economic criteria and premises for calculating product cost of designated hydrocarbons are also presented.

2.1 Natural Gas Composition

Natural gas was designated as the main feedstock. Since the study is not site specific, average composition of the pipeline gas available for the months of October and November 1983 from Trans Québec and Maritimes (Levesque 1983) has been used as a typical analysis for this study.

	mol %
Nitrogen	1.880
Carbon Dioxide	0.334
Methane	95.405
Ethane	2.217
Propane	0.151
i - Butane	0.007
n - Butane	0.006
	100.00

Higher heating value	
MJ/m^3	37.65
Btu/scf	994.0

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR TECHNOLOGY ASSESSMENT AND PROCESS SELECTION (Cont'd)

2.2 Product Criteria

As mentioned earlier, the main objective of the study is to investigate the feasibility of producing transportation fuel from natural gas. Gasoline was designated as a desired product. Hydrogen is to be recovered and sold as a by-product.

As a sub-objective of the study to calculate the preliminary product costs of a few selected hydrocarbons, the following cases were examined:

- Acetylene to be recovered as a main product for use as a potential feedstock for a wide variety of petrochemical products. Ethylene, propylene and butylene are the coproducts. Hydrogen is a by-product

- The recovered acetylene to be converted to ethylene as a feedstock for petrochemical production

2.3 Overall and Desired Product Efficiency

For technology assessment and process evaluation, overall and liquid product efficiencies have been used. In this study, a modified definition of product efficiency widely used for energy related processes has been considered. The modified definition of "overall" product efficiency is the total higher-heating-value (HHV) of all products divided by the total input energy of

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR TECHNOLOGY ASSESSMENT AND PROCESS SELECTION (Cont'd)

2.3 Overall and Desired Product Efficiency (Cont'd)

natural gas, LPG*, and electricity. For electricity, a thermal equivalent factor of 3600 kJ/kWh has been applied.

Since the objective of the study is to investigate electricity-based processes for the production of transportation fuel, the "liquid" product efficiency is also calculated. Here the term "liquid product" refers to gasoline.

2.4 Overall Plant Configuration, Plant Size and Battery Limits

All plants are considered as grass-root facilities with self-sufficient operation. Natural gas, LPG, electricity and water are major raw material and process inputs.

For transportation fuel case, the capacity of the designated "base" plant for each of the three designated reactor configurations is 12 500 bbl/d (1 987 350 litres/d) gasoline product. The capacity factor or onstream factor was assumed to be 8000 operating hours per year. The plant will be located in Québec.

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* LPG is used as a quench to stabilize the reactor product and also to recover part of the process heat

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR TECHNOLOGY ASSESSMENT AND PROCESS SELECTION (Cont'd)

2.4 Overall Plant Configuration, Plant Size and Battery Limits (Cont'd)

Make-up water is assumed to be available at the site. Steam is to be generated on site by recovering process heat and burning waste carbon black and fuel gas. Some electricity may be generated on site to utilize excess fuel gas. The remaining electricity required is to be purchased from the utility network.

For the hydrocarbon option, preliminary product cost calculations have been made for two plant sizes, 50 and 300 million lb/yr acetylene (22.68 and 136.116 million kg/yr). For the ethylene product, a plant capacity of 1 billion lb/yr (454 Gg/yr) has been considered.

2.5 Environmental Aspects

The conceptual plant design will meet all existing local, provincial and federal environmental regulations for particulates, liquid and gaseous effluents.

2.6 Economic Criteria

Preliminary or budget-type capital cost estimate, operating, maintenance and product cost calculations are made for each of the conceptual plants considered. Some of the basic financial assumptions and consistent set of criteria used for the study are summarized in Table 1-1. Details of the methodology and procedure followed are given in Chapter 4 entitled "Cost Considerations".

Table 2-1 Basic Financial Assumptions and Economic Criteria

Plant capacity	: 12 500 bbl/d (1 987 350 litre/d)
Plant location	: Québec
Constant Canadian dollars	: mid-1984 basis
Construction period	: 3 years
Capital investment schedule	

Year	%
1	22
2	48
3	30

Project and estimating contingency	: 10%
Financing	: 100% equity
Cost of capital	: 5% after taxes or about 9% before taxes
Tax rate	: 45%
Investment tax credit	: neglected
Plant operating efficiency	

Year	%
1	90
2-20	100

Plant life	: 20 years
Depreciation method	: straight line
Salvage value	: zero
Plant service factor	: 8000 hours/yr (0.91)
Natural gas cost	: $ 3.86/GJ ($4.0/$10^6$Btu)
LPG (propane/butane)	: $166/$m^3$ (16.6¢/litre)
Electricity cost	: 1¢/kWh
By-product credit for H_2	: $4.74 - $7.58/GJ ($5-8/10^6Btu)

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR TECHNOLOGY ASSESSMENT AND PROCESS SELECTION (Cont'd)

2.6 Economic Criteria (Cont'd)

All costs are given in "constant" mid-1984 Canadian dollars, therefore, escalation beyond mid-1984 is avoided. For the study, a 100% equity financing with cost of capital at 5% after tax or approximately 9% before tax was assumed*.

Cost estimates and data received for various process developers and licensors for different plant conditions, location and capacity have been adjusted on a consistent basis and converted for the plant location in Québec.

A 3-year construction period, 20-year plant operating life, straight-line depreciation and zero salvage value are assumed. The capital investment is based on a typical S-type curve and expenditure during the year is assumed to be uniform. The total capital investment includes the capital cost of onsite as well as necessary offsite support facilities. The total investment also includes allowance for contingency, prepaid license fees and royalties, initial catalyst and chemical costs, start-up expenses, land and working capital.

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* In "constant" dollar analysis, cost of capital is expressed in "real" terms i.e. current rates minus inflation (15 - 6 = 9%). Average inflation over 20 years was assumed to be 6% per year.

2.0 TECHNICAL AND ECONOMIC CRITERIA FOR TECHNOLOGY ASSESSMENT AND PROCESS SELECTION (Cont'd)

2.6 Economic Criteria (Cont'd)

Annual operating, maintenance and product costs for each of the plant configurations are reported for 8000 hours operation/year. However, it is assumed that the plant will operate at 90% efficiency during the first year and at full design capacity in the subsequent years.

Credit for by-product hydrogen is taken at $7.58/GJ ($8.00/10^6Btu). Benzene which is recovered as a by-product is not sold separately but is blended with the main product gasoline to improve its octane number. Impact on final product cost of gasoline considering lower by-product value at $4.74/GJ has also been examined.

For the hydrocarbon option, by-product credits have been taken for hydrogen at $4.74/GJ ($5/10^6Btu), benzene at $540/t (US $1.40/US gal.), ethylene at 55¢/kg (25¢/lb) and propylene/butylene at 42¢/kg (19¢/lb).

3.0 OVERALL PLANT CONFIGURATION

3.0 OVERALL PLANT CONFIGURATION

In this chapter, the overall conceptual plant design of a commercial liquid fuel plant (12 500 bbl/d gasoline product) is presented, using each of the three designated acetylene reactor configurations, namely the electric-arc, "preheated" hydrogen or hydrogen plasma, and the resistance-heated type reactors. The overall configuration includes the acetylene plant, catalytic hydrogenation, olefin oligomerization and hydrogen recovery units along with the required support systems, utility and offsite facilities to comprise an essentially integrated, self-sufficient operation.

The conceptual design presented here has been developed for the screening-type evaluation. Within the framework of budgetary limitations, SNC did not attempt to make a detailed or comprehensive analysis of the technologies, but instead based the design on the data supplied by various process developers. Data received were verified for consistency and accuracy. Where operating conditions were not specified by process developers, they were chosen by SNC based on the available information and criteria established. For example, operating conditions for the conceptual plant design using resistance-type reactor configuration were selected to maximize natural gas conversion and acetylene yield and minimize specific energy requirement (SER, kWh/kg of acetylene formed).

Natural gas, LPG and electricity are the basic raw material and process input, LPG being used as a quench to stabilize the intermediate acetylene and also to recover part of the process heat. Make-up water, catalysts and other chemicals are also needed but in relatively smaller quantities. During steady-state operation, steam is generated by recovering process heat and

burning fuel gas, residual oil and the carbon formed in the acetylene reactor. Excess fuel gas is used to generate part of the electricity, the balance is purchased from an external source.

The utility services include a full range of systems for a self-sufficient integrated operation of the plant such as steam generation, electricity and fuel gas distribution, cooling water, instrument air, boiler feedwater treatment, potable water, fire-water, storm and waste water facilities. Offsite requirements include such items as administration buildings, laboratory, maintenance building and warehouse, firehouse, in-plant road, fence, flares, onsite storage, etc.

In addition to conceptual design and process description, overall material and energy balance, product efficiencies and a utility summary are also given for each process. Also, a pertinent literature review related to the designated electricity-based acetylene processes and conversion to transportation fuel is presented. The review is not intended to be comprehensive but only to provide a perspective for a reader from which the results and conclusions presented in the report may be viewed.

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION

The overall block diagram showing the interrelationship between the various processing areas for an integrated liquid fuel plant is shown in Drawing 3-1. The main processing areas may be sub-divided as:

- Acetylene reactor/quenching
- Gas purification and primary product separation
- Acetylene hydrogenation
- Mobil Olefin to Gasoline and Diesel (MOGD) process, and
- Hydrogen recovery (Pressure Swing Absorption, PSA)

3.1.1 Acetylene Reactor and Quenching

Figure 3-1 is a recent design of the Huels electric-arc reactor. The gaseous feedstock enters the reactor tangentially through a turbulance swirl chamber R (700-800 mm diameter, 300-500 mm high) and passes with a rotary motion through the reaction zone A (diameter 85-105 mm, length approximately 1.5 m). The swirling motion of the downward flowing gas stream stabilizes the arc by essentially preventing upward convective movement of the hot gases which tends to break (or destabilize) the arc. Other process developers namely Dupont [Burk 1959], Westinghouse [Fey 1980], AVCO [Wragg 1980], etc. have opted to "rotate" the arc at high speed instead with the surrounding magnetic field.

The arc L burns between the bell-shaped cathode C and the tubular anode A (grounded at E). Both electrodes are hollow and are water-cooled. The striking point of the arc in the anode and/or cathode can be moved upward or downward, as required, and as such the arc is termed by Huels as "elastically suspended". The wall

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.1 Acetylene Reactor and Quenching (Cont'd)

thickness of the electrodes is 10-20 mm. With the exception of the insulator J all parts of the reactor are made of iron.

The arc is about 100 cm long and extends 40-50 cm into the anode. During the start-up operation, starting of the arc is achieved by an ignition electrode Z. The arc is operated with 8000 kW direct current at 7000 volts and a current of 1150 amperes. The power factor is nearly 0.75 and longitudinal field intensity is 70 V/cm. Other design features and details of the arc column are given in Appendix D.

As shown in Drawing 3-1, natural gas mixed with recycled gas is fed into one of several arc reactors. For the proposed plant capacity, 55 reactors each of 9 MW rating are required.

At Huels, other feedstocks such as refinery gas, liquid propane and butane, light and heavy naphthas have also been commercially tested. The recycled gas contains unreacted methane, some ethane, hydrogen, higher acetylenes, and dienes. Typically, the feed gas passes through the arc at about 1000 m/s and the residence time is in the order of 2 ms. Overall conversion of methane is about 70-75 wt%.

The hot gases containing mainly acetylene, hydrogen, higher acetylenes, dienes, unreacted methane and some carbon monoxide leaving the arc zone have substantial energy because of its high temperature (about 1500 C). About 20 cm below the anode A, cold propane/butane is introduced into the reactor through

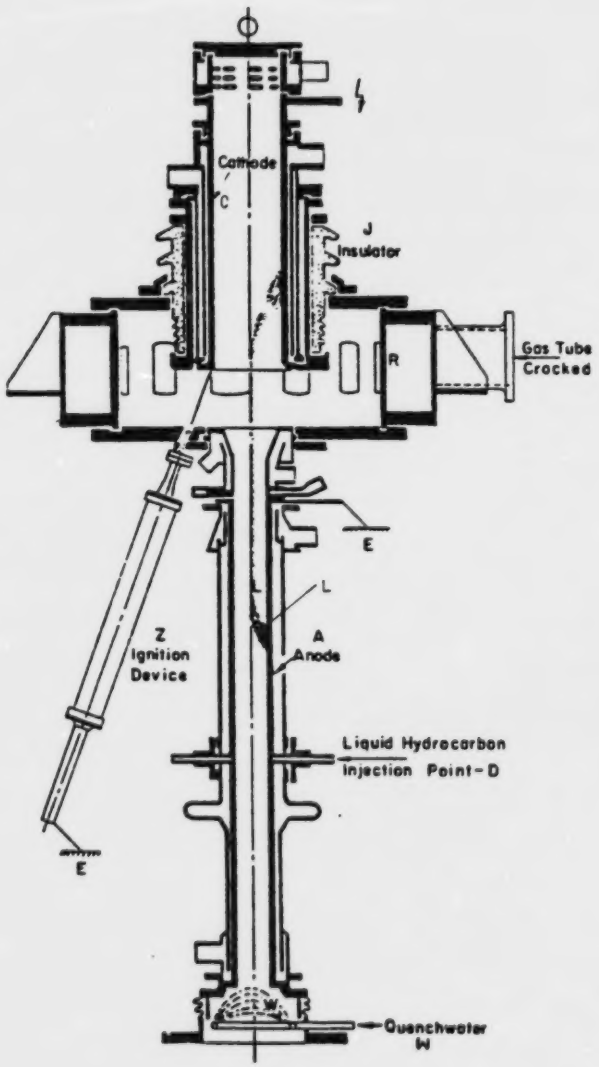

Figure 3-1 Schematic Diagram of the Huels Electric-Arc Reactor

Source: Muller, R. Chemische Werke Huels, Information Package prepared for SNC Inc., Ref 1050/08/Bu, December 23, 1983

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.1 Acetylene Reactor and Quenching (Cont'd)

several nozzles D to cool the gas stream to 800-1000 C and stabilize the acetylene formed. The cold LPG quench is mainly converted to ethylene. Immediately below the pipe A, the exiting hot reaction mixture is further cooled to about 200-250 C by means of a water spray W. The residence time during quenching is less than 2 ms.

Other methods of heat recovery investigated by Chemische Werke Huels such as use of boilers or its use in generating mechanical energy were not successful mainly because of longer residence time which reduced the desired product yield and caused considerable technical problems due to excessive carbon formation. This factor also militates against the use of a pebble cooler or checker-brick stove [Miller 1966] as this would also require the heat recovery to be carried out through complex operating cycles (cooling/quenching, regeneration to burn off carbon deposits, and purge). Therefore, for this study, the cold LPG quenching method was used.

The gas stream leaving the arc reactor has the following flow rate and composition:

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.1 Acetylene Reactor and Quenching (Cont'd)

	kg/h	mol %
C_2H_2*	47 627	13.8
C_2H_4	19 477	5.2
C_3/C_4 acetylenes	4 838	0.8
C_3/C_4 dienes	1 653	0.3
C_3H_6	4 382	0.8
C_4H_8	3 261	0.4
C_5H_6/C_6H_6	5 425	0.6
CH_4	25 005	11.8
C_2H_6	682	0.2
C_3H_8/C_4H_{10}	5 322	0.8
Higher hydrocarbons	2 564	0.2
H_2	15 029	56.6
CO	992	0.3
N_2	3 250	0.9
HCN	16	45 ppm
Carbon	11 709	7.3
	151 199	100.0

*acetylene formed in the arc zone = acetylene in the product stream
- acetylene in recycle feed
- acetylene formed during quenching

= 47 627 - 453 - 102 = 47072 kg/h

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.1 Acetylene Reactor and Quenching (Cont'd)

The reactor's specific energy requirement is about 11.3 kWh/kg of acetylene formed.

On a carbon-free basis, acetylene concentration in the product gas stream is about 15%. The existing gas stream undergoes a series of gas processing and product separation steps as described below:

3.1.2 Gas Purification and Primary Product Separation

The separation and processing of acetylene containing gas stream imposes certain restrictions and limitations due to the unstable and explosive behaviour of acetylene. Extensive physical data has been generated over the years concerning conditions which will decompose, explode or detonate acetylene. Also, experimental tests have been well documented by Miller [1966], Copenhaver and Bigeloro [1949] and by Kirk and Othmer [1964].

The results of decomposition and detonation studies on acetylene and its mixture with other gases indicate that operating conditions where the partial pressure of acetylene exceed 200 kPa (30 psi) should be avoided for safe operation. However, under special care acetylene has been transported and handled over long distances at pressures of several atmospheres [Copenhaver and Bigeloro 1949]. Similar limitations apply to operating temperatures which should not exceed 95-105 C. On the other hand, low temperatures lead to formation of liquid or solid-like acetylene and its homologs. The separation and purification of acetylene

3.1.2 **Gas Purification and Primary Product Separation** (Cont'd)

is further complicated by the presence of higher acetylenes which tend to polymerize easily.

In view of these severe operating restrictions a few commercial processes are available [Miller 1966, Tedeschi 1982] based on absorption-desorption techniques with one or more selective solvents namely anhydrous ammonia (SBA), cold methanol (Montecatini process), n-methylpyrrolidinone (BASF), di-methyl formamide (Dupont process), and water as in the original Huels process. At present, a modern separation process developed jointly by Linde AG [Kruis 1967] is in operation which is described below.

As shown in Drawing 3-1, most of the carbon black is initially removed from the product gas stream as dry carbon (3-4 wt% H_2O, 5 wt% hydrocarbons) by a series of cyclone separators. The balance is rejected as wet carbon (50 wt% H_2O, 25 wt% hydrocarbons) in a water spray towers. The wet carbon black is then sent to the horizontal filters to reject excess water. Both dry carbon and wet carbon filter cake is used as a fuel to generate steam for the process.

The solid-free gas (containing less than 3mg carbon/m^3 of gas) is treated in an oil venturi and scrubber to remove naphthalenes and other hydrocarbons. Benzene is further recovered as a by-product by distillation and blended with gasoline. The gas stream is then treated in a water scrubber to remove traces of hydrocyanic acid.

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.2 Gas Purification and Primary Product Separation (Cont'd)

The effluent gas is compressed to 1.5 MPa in four-stage reciprocating compressors and then treated in methanol scrubbers operating at 30 C to separate C_3/C_4 dienes and higher acetylenes before being recycled to the reactor.

The remaining gas stream is heated to about 150 C prior to feeding into the catalytic hydrogenation unit.

3.1.3 Acetylene Hydrogenation

Acetylene present in the gas stream is hydrogenated to olefins (mainly ethylene) since suitable processes for liquid fuel production are based mostly on olefin oligomerization rather than acetylene polymerization [Cooke and Ashraf 1984]. Also, acetylene polymerization processes are either at a developing stage (Allenger et al 1984] or have rapid catalyst deactivation problems [Tsai and Anderson, 1983].

Selective hydrogenation of acetylene to ethylene is carried out "in-situ" over Pd base catalyst at 150-175 C and 1.4 MPa using hydrogen present in the gas stream. Other typical catalysts include platinum, nickel and cobalt [Spiers 1963]. An extensive review on acetylene hydrogenation is described elsewhere (Miller 1966).

Conversion of acetylene to ethylene is high ($>$ 90%) and the reaction is highly exothermic (-175 kJ/gm-mole). Other side reactions which also take place are conversion of olefins such

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.3 Acetylene Hydrogenation (Cont'd)

as ethylene, propylene, and butylene to saturated hydrocarbons. Excess heat is used to preheat the exiting olefin-rich stream to 150 C and to generate steam for the process.

Some carbon deposition takes place on the catalyst and typically requires regeneration every six months. This is achieved by conventional steam/air decoking operation at 550 C. An auxiliary furnace is used to preheat the air and supply heat. The catalyst life is about 3-4 years.

The olefin-rich stream (52 wt% ethylene, 4 wt% propylene and 1.4 wt% butenes) is compressed to 3.45 MPa (500 psi) and preheated to 315 C (600 F) prior to feeding into the MOGD reactor.

3.1.4 Mobil Olefin to Gasoline and Distillates (MOGD) Process

MOGD is a catalytic process developed by Mobil Research and Development Corporation for converting light olefenic compounds to high quality gasoline and/or distillates. The process is based on Mobils' synthetic zeolite catalyst (ZSM-5) which selectively oligomerizes the light olefins to the desired product. Though normally designed to process propylene and butylene, MOGD is applicable to a wide range of feedstreams ranging from ethylene to olefinic naphtha [Tabak 1984].

The process is also flexible with respect to the final product so as to produce distillate/gasoline ratios ranging from 0/100 to

3.1.4 Mobil Olefin to Gasoline and Distillates (MOGD) Process (Cont'd)

90/100. Based on the analysis of feedstream supplied by SNC to Mobil, the process developer recommended that gasoline production would be desirable [Weiss 1984]. Suggested reactor conditions are 250-315 C and pressure of 1.8-3.4 MPa. The product gasoline has high octane number (92 RON, 79 MON) and meets most of the standard specifications. API gravity is 62.

The reactor is highly exothermic necessitating heat control to avoid temperature runaway problems. Four fixed-bed reactors are used, three on line and one in generation. The three on line reactors are in series with inter-reactor coolers and liquid recycle to control the heat. The reactor with the oldest catalyst is in the first position whereas the reactor with fresh catalyst is in the third or the last position, thus a descending temperature profile is maintained. Since high molecular weight product is thermodynamically favored by low temperature, this combination of low temperature with high activity catalyst in the last reactor is directionally consistent for maximum selectivity and conversion to the desired gasoline product.

The reactor effluent is separated into three streams by conventional flash separation and distillation methods. The offgas stream containing methane and hydrogen is sent to the Pressure Swing Adsorption (PSA) unit for byproduct hydrogen recovery. The LPG stream is recycled to the acetylene reactor for quenching. The gasoline product is blended with the byproduct benzene and sent to storage.

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.4 Mobil Olefin to Gasoline and Distillates (MOGD) Process (Cont'd)

Flow rate and composition of these streams are given below:

	Offgas		LPG		Gasoline
Flow rate, kg/h	43 085		19 640		60 278
Composition, wt%					
H_2	26.0	C_3H_6	12.9	Gasoline	100.0
CH_4	57.8	C_3H_8/C_4H_{10}	72.8		100.0
C_2H_6	6.4	C_4H_8	14.3		
CO	2.2		100.0		
N_2	7.6				
	100.0				

3.1.5 Hydrogen Recovery Using Pressure Swing Adsorption (PSA) Unit

The offgas leaving from MOGD process at 40 C is compressed from 1.24 MPa (180 psi) to 2.4 MPa (350 psi) and fed into the Union Carbides' Polybed Pressured Swing Adsorption unit. Essentially the PSA unit is based on the principle of pressure reduction to remove adsorbed components from the molecular sieve (adsorbent) which adsorbs nearly all the components other than hydrogen from the inlet gas stream.

A complete PSA system cycle consists of the following four basic steps: adsorption, depressurization, purge at low pressure, and repressurization. Typically when the reactor 1 is on the adsorption step providing high-purity hydrogen at high-pressure, adsorber 2 is in the depressurization phase providing gas for

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.5 Hydrogen Recovery Using Pressure Swing Adsorption (PSA) Unit (Cont'd)

purging of adsorber 3 and the repressurization of adsorber 4. Adsorber 3 is being purged at low pressure (14 kPa) for removal of other components in the tail-gas line. Adsorber 4 is repressurized by the gas from adsorber 2, and the gas is then recovered when this vessel comes on line.

The pressure swing cycle can be accomplished by using any number of adsorbent beds. In its simplest form, only two adsorbent beds are needed; one being on adsorption while the other bed is undergoing the three steps of regeneration. The conventional two-bed system results in the lowest equipment cost but has a major disadvantage in allowing rather large unrecoverable hydrogen losses during the repressurization step. In addition, the two-bed system is limited to relatively low throughputs.

The four-bed HYSIV^{R} PSA unit minimizes hydrogen losses occuring during depressurization. However, as a general rule, use of these units is restricted to 12 MM scf/d (13 400 Nm^3/h) of hydrogen product capacity for optimum overall economics taking into consideration the equipment cost and hydrogen recovery. For greater capacities, polybed PSA units which incorporate a great number of beds and cycle modifications to further increase hydrogen recovery and reduce adsorbent inventory are advantageous. Capacities as high as 70 MM scf/d (78 275 Nm^3/h) of hydrogen can be achieved in a single train of PSA units incorporating shop-fabricated vessels. For the study, two trains each of 55 MM scf/d (61 500 Nm^3/h) of hydrogen capacity have been considered.

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.5 Hydrogen Recovery Using Pressure Swing Adsorption (PSA) Unit (Cont'd)

The recovered by-product hydrogen 99.4 wt% pure. Impurities include 0.6% nitrogen and 10 ppm carbon monoxide. Overall hydrogen recovery is about 90%.

The waste gas stream containing mainly methane, some ethane, carbon monoxide, hydrogen and nitrogen is used partly as inplant fuel (100 GJ/h). The remaining portion is used as a stripping gas and the balance is recycled directly to the acetylene reactor.

3.1.6 Utilities and Offsite Facilities*

The utility services include a full range of systems necessary for the operation of the plant, such as raw water treatment, cooling water and boiler feed water, condensate treatment, steam and electricity distribution, plant and instrument air system, fuel gas purification and distribution.

River water is assumed to be the source of make-up water for the plant. This water may be treated in a cold lime unit and filtered to reduce hardness, carbonate alkalinity, turbidity and suspended organics. The softened water is used directly as cooling tower make-up water and as process water. It is also

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* not shown on Drawing 3-1

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.6 Utilities and Offsites Facilities (Cont'd)

supplied to the demineralization unit where cation, anion, and mixed-bed ion-exchangers are used for boiler feed water (BFW) preparation. The demineralized water is sent for deaeration and then distributed to onsite and offsite steam generators.

Waste water from the effluent control section, drained-off water from product storage areas, and contaminated rain water are collected and treated in an API separator for removal of oil contamination before passing to the intermediate treatment section. Cooling tower blowdown water is also sent for intermediate treatment where the water is neutralized and subjected to chemical flocculation and dissolved air flotation. Part of the treated water is used as make-up for the cooling towers, and the excess passes to the outfall system. Uncontaminated rainwater is also drained to the outfall system.

All the steam requirements for the process are met by recovering process heat and burning residual oil, carbon black and excess fuel gas. Some electricity is also generated on site.

Other offsite requirements include intermediate and final product storage, flare system, laboratories, administration buildings, warehouses, firehouse, onsite roads and shipment facilities.

3.1.7 Overall Material and Energy Balance, Product Efficiencies and Utilities Summary

The interrelationship between various processing areas and the material balance of a conceptual, commercial, integrated liquid

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.7 Overall Material and Energy Balance, Product Efficiencies and Utilities Summary (Cont'd)

fuel plant is presented in Drawing 3-1. The overall material balance summary is presented in Table 3-1. The utilities balance, i.e. the consumption and production of steam, fuel gas, etc. are summarized in Table 3-2.

The main product is gasoline. Hydrogen is the byproduct. Residual oil, carbon black and excess fuel gas are used to generate steam and 95 MW electricity within the plant. All the low-pressure steam required for the process (32 675 kg/h) is generated by recovering process heat from the acetylene hydrogenation unit. The excess LP steam (62 240 - 32 675 = 29 565 kg/h) is used for generating high-pressure steam.

The make-up water requirement is 910 m^3/h, or about 4000 US GPM. The large volume of this make-up water is mainly due to losses and consumption of water in water-coolers.

The "overall product" efficiency is 60.3%. The definition used in the total heating value of all products (gasoline and hydrogen) divided by the input energy of natural gas, LPG and electricity. Similarly, the desired gasoline product efficiency is 41.8%.

Input	Flowrate	HHV kJ/kg	Energy GJ/h
Natural Gas	81 170 kg/h	52 515	4262.6
LPG	14 225 kg/h	51 400	731.2
Electricity	532 MW	3600 kJ/kWh	1915.2
			6909.0

3.1 LIQUID FUEL PLANT WITH ELECTRIC-ARC ACETYLENE REACTOR CONFIGURATION (Cont'd)

3.1.7 Overall Material and Energy Balance, Product Efficiencies and Utilities Summary (Cont'd)

Output	Flowrate	HHV kJ/kg	Energy GJ/h
Gasoline	62 339 kg/h	46 272	2884.8
LPG	10 816 kg/h	118 700	1283.9
			4168.7

Table 3-1 Overall Material and Energy Balance Summary
Plant Capacity: 12 500 bbl/d Gasoline Product
Reactor Configuration: Electric-Arc (Chemische Werke Huels)

	Flowrate/Quantity	Unit Basis*** per kg of gasoline
Process Input		
Natural Gas	81 170 kg/h	1.30 kg
LPG	14 225	0.23
Scrubbing Oil	930	-
Methanol	465	-
Electricity		
Acetylene Reactor	532 MW	8.5 kWh
*Plant Use	95	1.5
	627	10.0
Utilities Required		
*HP Steam (2 MPa)	62 550 kg/h	1.0 kg
**LP Steam (0.4 MPa)	32 675	-
Cooling Water (t_e=26C, Δ t=10C)	29 375 m^3/h	0.47 m^3
Make-up Water	910	0.014
Instrument Air (0.5 MPa, Dew Point = -25C)	1 200 Nm^3/h	-
Nitrogen (0.4 MPa)	975	-
Products/Byproducts		
Gasoline (12 500 bbl/d)	62 339 kg/h	-
Hydrogen	10 816	0.17 kg
*Residual	4 030	0.065
*Carbon black (dry)	11 245	0.18
(wet)	5 623	0.09
*Fuel Gas (110 GJ/h)	3 630	1.76 MJ

* Residual oil, carbon black and fuel gas are used to generate steam and some electricity for inplant use.

** LP steam requirement (32 675 kg/h) is met by steam produced (62 240 kg/h) in the acetylene hydrogenation unit. Excess LP steam (62240 - 32675 = 29565 kg/h) is used for generating HP steam. See Table 3-2 for utilities summary.

*** 1 kg of gasoline = 1.3342 litres.

Table 3-2 Utilities Summary for a 12 500 bbl/d Gasoline Plant Using Electric-Arc Reactor (Chemische Werke Huels)

	Residual Oil	Carbon Black (dry)	Carbon Black (wet)	Fuel Gas	Electricity	Steam HP(2MPa)	Steam LP(0.4 MPa)	Cooling Water	Make-up H_2O	Instrument Air	Nitrogen
Process Areas	kg/h	kg/h	kg/h	GJ/h	MW	kg/h	kg/h	m^3/h	m^3/h	Nm^3/h	Nm^3/h
Acetylene Unit	-	-	-	-	586*	60685	32675	28005	-	700	700
Hydrogenation Unit	-	-	-	-	3	1865	-	1245	-	235	-
Mobil Process	-	-	-	110	33	-	-	125	-	70	-
PSA Units	-	-	-	-	5	-	-	-	-	115	-
Offsites	4030	11245	5623	-	*	-	29565**	-	-	80	275
Total Consumption	4030	11245	5623	110	627	62550	62240	29375	-	1200	975
Acetylene Unit	4030	11245	5623	-	-	-	-	-	-	-	-
Hydrogenation Unit	-	-	-	-	-	-	62240	-	-	-	-
Mobil Process	-	-	-	-	-	-	-	-	-	-	-
PSA Unit	-	-	-	-	-	-	-	-	-	-	-
Offsites	-	-	-	110	95	-	-	29375	910	1200	975
Total Generation	4030	11245	5623	110	95	62550	62240	29375	910	1200	975

* Offsite requirements are included.
** Excess LP steam (62240 - 32675 = 29565 kg/h) is used for generating HP steam.

* Offsite requirements are included.
** Excess LP steam (62240 - 32675 = 29565 kg/h) is used for generating HP steam.

3.2 LIQUID FUEL PLANT WITH HYDROGEN-PLASMA ACETYLENE REACTOR

The overall processing scheme is similar to that described earlier for the electric-arc reactor except that:

- the hydrogen plasma reactor is used for acetylene production
- a portion of recovered hydrogen is recycled to the acetylene reactor to preheat the feed

Differences in the operating conditions and product yields are described below in the text. Design features of the hydrogen plasma reactor system are also presented.

3.2.1 Hydrogen Plasma Acetylene Reactor

The use of hydrogen as an auxiliary gas for heating and pyrolysis of natural gas is particularly attractive as it acts as a suppressant for carbon formation by reducing acetylene decomposition [Kanugi 1961]. Also, hydrogen is formed as a byproduct with the acetylene, therefore, use of hydrogen does not complicate the product separation and recovery procedure.

The reactor $H_2 \longrightarrow 2H$ in the arc zone is highly endothermic (ΔH=429 kJ/gm-mole). At the plasma temperature of 5000 C, the concentration of hydrogen ions is almost 99%, dropping to 1% at 2500 C. Therefore, on cooling from the plasma state to acetylene reaction temperature ($\simeq$ 1500 C), it provides 565 kJ/mol. Also, heat content of hydrogen gas leaving the arc zone at 1200 C is only 5% of its enthalpy at 5000 C, thus the penalty for heat loss is low.

3.2 LIQUID FUEL PLANT WITH HYDROGEN-PLASMA ACETYLENE REACTOR (Cont'd)

3.2.1 Hydrogen Plasma Acetylene Reactor (Cont'd)

The hydrogen-plasma reactor has been used by Westinghouse [Fey and Reed 1980] for production of acetylenes from hydrocarbons including natural gas, from coal by AVCO [AVCO 1981] and by Chemische Werke Huels [Muller and Peukert 1983] and others. Essentially, the concept is similar, however distinguishing features of the Westinghouse plasma acetylene reactor system are described below.

As shown in Figure 3-2, the plasma torch consists of a closely-spaced pair of tubular, water-cooled copper electrodes within which the electric arc is magnetically rotated at high speed (1000-3000 revolutions/sec) by internally mounted solenoid coils. During operation, the feed gas is injected through the narrow gap between the electrodes which are spaced approximately 1 mm apart. A source of 4-6 kV electric power is typically maintained to provide a spark over in the electrode gap, initiating the arc discharge, which is blown or carried into the arc zone by the incoming feed gas stream.

The patented spark over feature provides a design which is self-starting when energized and self-stabilizing during operation. In fact, this distinguishing feature essentially allows the AC operation without the use of auxiliary means. However, recent developments by Westinghouse favor the use of a DC-type plasma system. Several DC torches of different power ratings ranging from small scale 75 kW unit (Marc 3) to commercial-scale (Marc 100) are being offered. Other design features of the

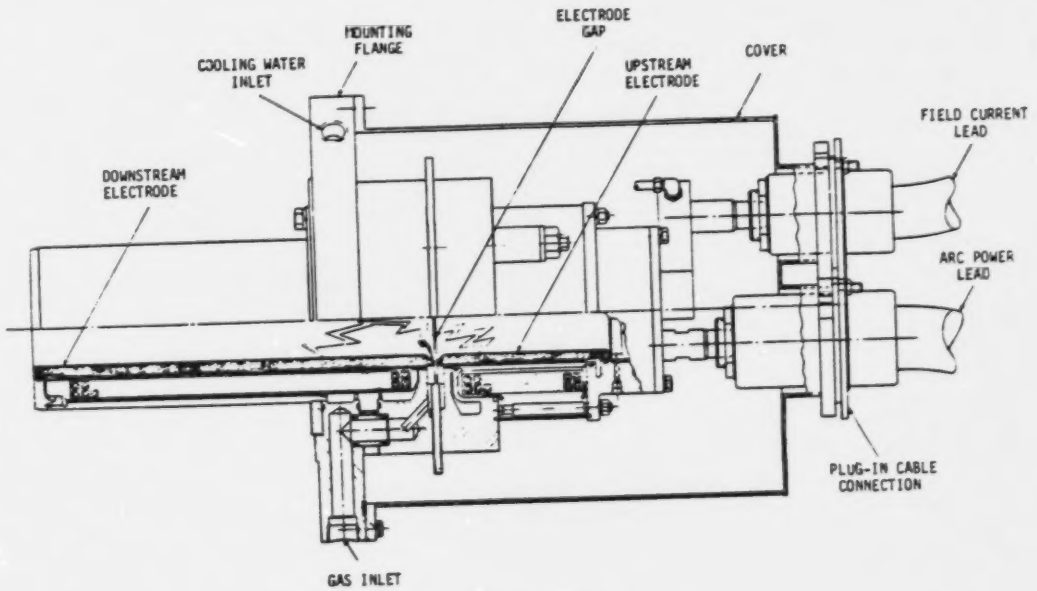


Figure 3-2: Schematic Diagram of Westinghouse Plasma Reactor

Table 3-3 Design Features of Westinghouse Plasma Reactor (Marc 100)

Arc column	:	non-transferred, magnetic arc rotation, water-cooled, tubular electrodes. typical dimensions 1 m long, 60 cm overall dia., electrode bore 10 cm dia.
Operating characteristics	:	arc voltage 2.5 kV DC, arc current approx. 3 kA DC for reducing gas
Torch rating	:	8-10 MW
Maximum gas flowrate	:	2000 Nm^3/h
Efficiency	:	80-85%
Typical electrode life	:	300-400 h
Cooling water requirement	:	40-50 m^3/h
Power supply requirement	:	DC thyristor type supply, 4-6 kV AC input, includes isolation transformer, smoothing reactor, thyristor converters, and control circuit
Control and instrumentation:		series of interlocked controls to both interface and control plasma system operation, also provides auto shutdown if key designated parameters of subsystems fails (loss of gas flow or water flow etc.), can be interfaced with data logger and/or process computer

Source: Reed [1984]

3.2 LIQUID FUEL PLANT WITH HYDROGEN-PLASMA ACETYLENE REACTOR (Cont'd)

3.2.1 Hydrogen Plasma Acetylene Reactor (Cont'd)

Marc 100 plasma unit (under development) are summarized in Table 3-3.

Temperature in the arc zone can be adjusted by varying the arc current independently of the gas feedstock flowrate, and thereby providing higher efficiency. Also, the combination of large throughput flowrates possible for the feedgas and high arc rotation speeds produces a highly turbulent mixing action and yields a high operating efficiency (typically 75-85%).

The copper electrodes are water cooled. Typically, electrodes are replaced after every 300-400 hours of operation. Installation arrangements are designed such that an electrode change can be performed in less than one hour. This is normally accomplished by separating the head from the base and replacing the downstream and upstream electrodes which are mounted on axially sliding O-ring seals.

For the 12 500 bbl/d liquid fuel plant, about 70 plasma torches are required. As shown in Drawing 3-1, natural gas mixed with recycle stream is fed into the hydrogen plasma reactor. The hot product gas is quenched with cold LPG and water as in the case of the electric-arc reactor described earlier. The flowrate and composition of the gas leaving the reactor is as follows:

3.2 LIQUID FUEL PLANT WITH HYDROGEN-PLASMA ACETYLENE REACTOR (Cont'd)

3.2.1 Hydrogen Plasma Acetylene Reactor (Cont'd)

	kg/h	mol %
C_2H_2*	48 116	12.1
C_2H_4	19 074	4.4
C_3/C_4 acetylenes	3 510	0.5
C_3/C_4 dienes	1 163	0.2
C_3H_6	4 221	0.7
C_4H_8	708	0.1
C_5H_6/C_6H_6	4 909	0.4
CH_4	22 601	9.2
C_2H_6	111	0.03
C_3H_8/C_4H_{10}	4 867	0.6
Higher hydrocarbons	2 323	0.1
H_2	20 838	68.0
CO	876	0.2
N_2	2 869	0.7
HCN	14	32 ppm
H_2O	2 878	1.2
Carbon	4 936	1.6
	144 014	100.0

The reactor SER is about 14.5 kWh/kg of acetylene formed*. On a carbon-free basis, the acetylene composition is about 12.3 mol %. The lower concentration is primarily due to some dilution of the product gas stream by preheated hydrogen.

The reactor effluent undergoes a similar gas processing and product separation steps as described earlier.

---

*acetylene formed in the arc zone = acetylene in the product stream
- acetylene in recycle feed
- acetylene formed during quenching

= 48 116 - 460 - 97 = 47 559 kg/h

3.2 LIQUID FUEL PLANT WITH HYDROGEN-PLASMA ACETYLENE REACTOR (Cont'd)

3.2.2 Overall Material and Energy Balance, Product Efficiencies and Utilities Summary

The overall material and utilities balance summary is given in Tables 3-5 and 3-6 respectively.

All the low-pressure steam required for the process is generated by recovering process heat. Excess LP steam is used for generating HP steam.

The make-up water requirement is about 705 m^3/h or about 3100 US gpm.

The "overall product" efficiency is 54.2 % as calculated below. Similarly, the desired gasoline product efficiency is 41.0 %.

Input	Flow rate	HHV kJ/kg	Energy GJ/h
Natural Gas	74 403 kg/h	52 515	3907.3
LPG	12 612 kg/h	51 400	648.3
Electricity	690 MW	3 600 kJ/kWh	2484.0
			7039.6
Output			
Gasoline	62 386 kg/h	46 276	2887.0
Hydrogen	7 850 kg/h	118 700	931.8
			3818.8

Table 3-4 Overall Material Balance and Energy Balance
Plant Capacity: 12 500 bbl/d Gasoline Product
Reactor Configuration: Preheated Hydrogen or Hydrogen Plasma Reactor

	Flowrate/Quantity	Unit Basis*** per kg of gasoline
Process Input		
Natural Gas	74 403 kg/h	1.19 kg
LPG	12 612	0.20
Scrubbing Oil	948	-
Methanol	472	-
Electricity		
Acetylene Reactor	690 MW	11.1 kWh
*Plant Use	90	1.4
	780	12.5
Utilities Required		
*HP Steam (2 MPa)	57 530 kg/h	0.92 kg
**LP Steam (0.4 MPa)	27 800	-
Cooling Water (t_e=26C, Δt=10C)	22 675 m^3/h	0.36 m^3
Make-up Water	705	0.011 m^3
Instrument Air (0.5 MPa, Dew Point = -25C)	990 Nm^3/h	-
Nitrogen (0.4 MPa)	755	-
Products/Byproducts		
Gasoline (12 500 bbl/d)	62 386 kg/h	-
Hydrogen	7 850	0.13 kg
*Residual	3 933	0.063
*Carbon black (dry)	4 936	0.079
(wet)	2 843	0.045
*Fuel Gas (315 GJ/h)	5 510	5.0 MJ

* Residual oil, carbon black and fuel gas are used to generate steam and some electricity for inplant use.

** LP steam requirement (27 800 kg/h) is met by steam produced (62 880 kg/h) in the acetylene hydrogenation unit. Excess LP steam (62880 - 27880 = 35080 kg/h) is used for generating HP steam. See Table 3-4 for utilities summary.

*** 1 kg of gasoline = 1.3342 litres.

Table 3-5 Utilities Summary for a 12 500 bbl/d Gasoline Plant Using Hydrogen Plasma Reactor Configuration

	Residual oil	Carbon Black (dry)	Carbon Black (wet)	Fuel Gas	Electricity	Steam HP(2MPa)	Steam LP(0.4 MPa)	Cooling Water	Make-up H_2O	Instrument Air	Nitrogen
Process Areas	kg/h	kg/h	kg/h	GJ/h	MW	kg/h	kg/h	m^3/h	m^3/h	Nm^3/h	Nm^3/h
Acetylene Unit	-	-	-	-	737*	55600	27800	21300	-	455	455
Hydrogenation Unit	-	-	-	-	3	1930	-	1250	-	240	-
Mobil Process	-	-	-	113	34	-	-	125	-	70	-
PSA Units	-	-	-	-	6	-	-	-	-	130	-
Offsites	3933	4936	2843	202	*	-	35080**	-	-	95	300
Total Consumption	3933	4936	2843	315	780	57530	62880	22675	-	990	755
Acetylene Unit	3933	4936	2843	-	-	-	-	-	-	-	-
Hydrogenation Unit	-	-	-	-	-	-	62880	-	-	-	-
Mobil Process	-	-	-	-	-	-	-	-	-	-	-
PSA Unit	-	-	-	-	-	-	-	-	-	-	-
Offsites	-	-	-	315	90	57530	-	22675	705	990	755
Total Generation	3933	4936	2843	315	90	57530	62880	22675	705	990	755

* Offsite requirements are included.
** Excess LP steam (62880-27800 = 35080 kg/h) is used for generating HP steam.

* Offsite requirements are included.
** Excess LP steam (62880 - 27880 = 35080 kg/h) is used for generating HP steam.

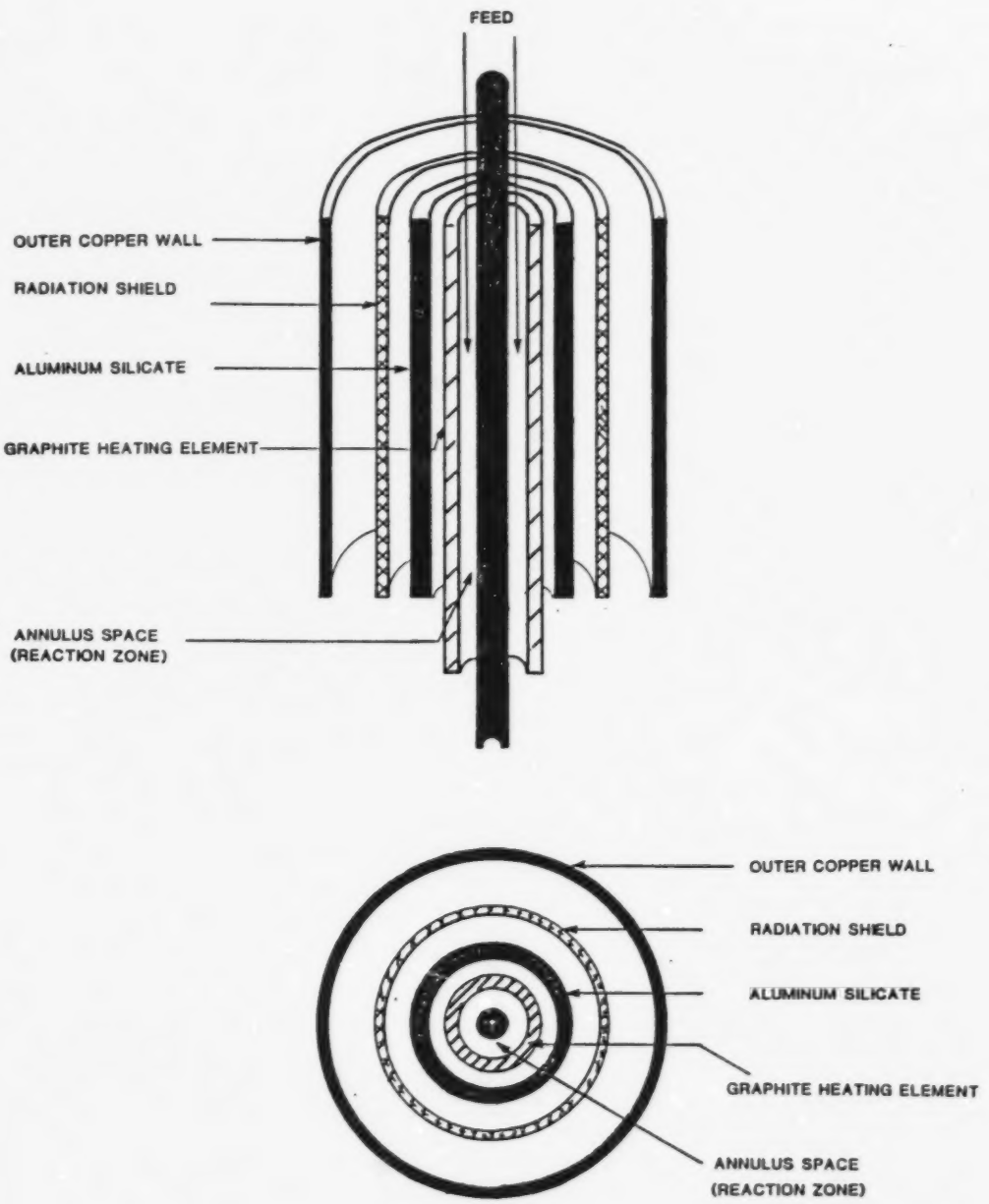


Figure 3-3: Schematic Diagram of Resistance-Heated Reactor

3.3 LIQUID FUEL PLANT WITH RESISTANCE TYPE ACETYLENE REACTOR

The overall processing scheme and material balance is presented in Drawing 3-1. The design features and description of the resistance type acetylene reactor configuration are given below.

3.3.1 Resistance Type Acetylene Reactor Configuration

The resistance type acetylene reactor consists of a system of vertically mounted cylindrical tubes which are progressively larger in diameter, all contained within an outermost metal tube.

Figure 3-3 is a simplified cross-sectional and axial sketch of the resistance type bench-scale reactor proposed by Happel and Kramer [1964]. The narrow annulus space (1 cm o.d, 0.5 cm i.d., and 12 cm length) serves as a reaction zone. A graphite resistor (type ATJ, National Carbon Co.) was used as a heating element. The radiation shield is used to minimize heat losses. The outer shell of the reactor including the 1-cm thick side plate is constucted of copper. All the outer walls are water cooled.

The main advantage of the "isothermal" resistance type reactor (which is essentially an isothermal-wall reactor) over the electric-arc/plasma processes is that the electrically heated reaction tube is easier to control and maintain a uniform temperature profile, thus making it possible to obtain higher yields of acetylene. An experimental study by Happel and Kramer [1967] showed that methane conversion up to 95% with acetylene yield up to 90% is achievable. The desired operating condition is reaction temperatures in the range of 1600-1800 C with residence time of less than 3 ms and pressures between 30-60 kPa.

3.3 LIQUID FUEL PLANT WITH RESISTANCE TYPE ACETYLENE REACTOR (Cont'd)

3.3.1 Resistance Type Acetylene Reactor Configuration (Cont'd)

Further study with a similar multi-tubular reactor having a capacity 50 times higher was built and operated by SINTEF [Helgesen et al 1967], although several modifications were made [Holmen et al 1976]. The reactor tube (CGW graphite from Union Carbide) used as a heating element was 60 cm long having 2-cm outer diameter. Two different inside diameters were tested, 0.7 cm and 1 cm. A protective lining of aluminum was used*. Undesirable carbon deposits were removed periodically by controlled air combustion. The capacity of the PDU used was 1 MW.

For the study, the multi-tubular resistance reactor of 1-m length and a 10 MW power rating was considered. For the 12 500 bbl/d capacity, about 55 such reactors are required. The selected operating conditions are reaction temperature of 1750-1775 C, residence time of 2 ms and pressure of 50 kPa. These are based on the analysis of bench-scale and pilot plant data supplied by the process developer and chosen not only to maximize methane conversion and acetylene yield but also to minimize the reactor SER (kWh/kg of acetylene produced).

A conservative value of 88% for methane conversion was assumed. Acetylene yield of 80% was taken. Product yield and distribution data reported for methane only was extended for other components

---

* Different materials of construction (e.g. SiC, Mo etc.) were considered in addition to two types of reactor configuration (rod and tubular). The tubular reactor with carbon as a heating element and alumina lining was found to be more flexible and reliable.

3.3 LIQUID FUEL PLANT WITH RESISTANCE TYPE ACETYLENE REACTOR (Cont'd)

3.3.1 Resistance Type Acetylene Reactor Configuration (Cont'd)

present in the natural gas feed stream. Also, adjustments were made for LPG quench, since in the experimental runs, cold product gas was used as a quenchant.

The flowrate and composition of the product gas stream leaving the resistance reactor are as follows:

	kg/h	mol %
C_2H_2*	56 184	17.5
C_2H_4	12 847	3.7
C_3/C_4 acetylenes	3 401	0.6
C_3/C_4 dienes	270	0.05
C_3H_6	4 377	0.85
C_4H_8	575	0.08
C_5H_6/C_6H_6	732	0.08
CH_4	16 051	8.1
C_2H_6	1 369	0.4
C_3H_8/C_4H_{10}	2 121	0.35
Higher hydrocarbons	1 555	0.13
H_2	15 082	61.1
CO	1 187	0.35
N_2	2 714	0.8
HCN	13	40 ppm
Carbon	6 717	1.4
	128 392	100.0

The reactor effluent is further processed to gasoline product as described earlier.

---

*acetylene formed in the arc zone = acetylene in the product stream
- acetylene in recycle feed
- acetylene formed during quenching

= 56 184 - 537 - 105 = 55 542 kg/h

3.3 LIQUID FUEL PLANT WITH RESISTANCE TYPE ACETYLENE REACTOR (Cont'd)

3.3.2 Overall Material and Energy Balance, Product Efficiencies and Utilities Summary

A summary of overall material and utilities balances are given in Tables 3-6 and 3-7 respectively.

The make-up water requirement is 815m^3/h or about 4740 US gpm.

The "overall product" efficiency is 61.7%. Similarly, the desired gasoline product efficiency is 43.6%.

Input	Flow rate	HHV kJ/kg	Energy GJ/h
Natural Gas	73 042 kg/h	52 515	3 835.8
LPG	15 285 kg/h	51 400	785.6
Electricity	548 MW	3 600 kJ/kWh	1 972.8
			6 594.2
Output			
Gasoline	62 139 kg/h	46 276	2 875.5
Hydrogen	10 034 kg/h	118 700	1 191.0
			4 066.5

Table 3-6 Overall Material and Energy Balance Summary
Plant Capacity: 12 500 bbl/d Gasoline Product
Reactor Configuration: Resistance Heated Reactor

	Flowrate/Quantity	Unit Basis*** per kg of gasoline
Process Input		
Natural Gas	73 042 kg/h	1.17 kg
LPG	15 285	0.25
Scrubbing Oil	820	-
Methanol	420	-
Electricity		
Acetylene Reactor	548 MW	8.8 kWh
*Plant Use	86	1.4
	634	10.2
Utilities Required		
*HP Steam (2 MPa)	63 250 kg/h	1.02 kg
**LP Steam (0.4 MPa)	32 330	-
Cooling Water (t_e=26C, Δ t=10C)	26 115 m^3/h	0.42 m^3
Make-up Water	815	0.013 m^3
Instrument Air (0.5 MPa, Dew Point = -25C)	905 Nm^3/h	-
Nitrogen (0.4 MPa)	735	-
Products/Byproducts		
Gasoline (12 500 bbl/d)	62 139 kg/h	-
Hydrogen	10 034	0.16 kg
*Residual	1 090	0.018
*Carbon black (dry)	6 717	0.11
(wet)	3 870	0.06
*Fuel Gas (305 GJ/h)	7 020	5.0 MJ

* Residual oil, carbon black and fuel gas are used to generate steam and some electricity for inplant use.

** LP steam requirement (32 330 kg/h) is met by steam produced (73 415 kg/h) in the acetylene hydrogenation unit. Excess LP steam (73 415 - 32 330 = 41 085 kg/h) is used for generating HP steam. See Table 3-7 for utilities summary.

*** 1 kg of gasoline = 1.3342 litres.

Table 3-7 Utilities Summary for a 12 500 bbl/d Gasoline Plant Using Resistance - Heated Reactor Configuration

	Residual oil	Carbon Black (dry)	Carbon Black (wet)	Fuel Gas	Electricity	Steam HP(2MPa)	Steam LP(0.4 MPa)	Cooling Water	Make-up H_2O	Instrument Air	Nitrogen
Process Areas	kg/h	kg/h	kg/h	GJ/h	MW	kg/h	kg/h	m^3/h	m^3/h	Nm^3/h	Nm^3/h
Acetylene Unit	-	-	-	-	595*	64950	32330	24725	-	440	440
Hydrogenation Unit	-	-	-	-	3	1700	-	1280	-	210	-
Mobil Process	-	-	-	100	31	-	-	110	-	65	-
PSA Units	-	-	-	-	5	-	-	-	-	95	-
Offsites	1090	6717	3870	205	*	-	41085**	-	-	95	295
Total Consumption	1090	6717	3870	305	634	66650	73415	26115	-	905	735
Acetylene Unit	1090	6717	3870	-	-	-	-	-	-	-	-
Hydrogenation Unit	-	-	-	-	-	-	73415	-	-	-	-
Mobil Process	-	-	-	-	-	-	-	-	-	-	-
PSA Unit	-	-	-	-	-	-	-	-	-	-	-
Offsites	-	-	-	305	86	66650	-	26115	815	905	735
Total Generation	1090	6717	3870	305	86	66650	73415	26115	815	905	735

* Offsite requirements are included.
** Excess LP steam (73415-32330 = 41085 kg/h) is used to generate HP steam.

* Offsite requirements are included

** Excess LP Stream (73 415 - 32 330 = 41 085 kg/h) is used to generate HP steam

3.4 OVERALL PLANT CONFIGURATION FOR HYDROCARBON PRODUCTION

A simplified flow diagram for acetylene production using electric-arc processing Chemische Werke Huels is given Figure 3-4. The processing scheme consists of acetylene reactor system, carbon separation and recovery unit, and series of scrubbers to remove impurities and higher acetylenes as described earlier in Section 3.1.1 (Acetylene Reactor and Quenching) and Section 3.1.2 (Gas Purification and Primary Product Separation).

The product gas leaving the methanol scrubber undergoes further processing steps as follows:

- Separation of hydrocarbons and higher olefins (C_3^-) using an octane scrubbing unit. Propane and butane are recovered in the subsequent low-temperature fractionation units as by-products, the remaining hydrocarbon containing gas is recycled to the acetylene reactor.

- Recovery and separation of acetylene (99% pure) in the methanol/n-methylpyrrilidone scrubbing unit.

- Separation of the residual gas into hydrogen, ethylene, methane and fuel gas stream. Hydrogen and ethylene are recovered as by-products. Methane is recycled to the acetylene arc reactor. Fuel gas is used for steam and power generation.

The overall material balance summary for the two plant capacities (300 and 50 million lb/yr acetylene) is given in Appendix B. The overall, desired products (acetylene and ethylene) and specific product (acetylene) efficiencies have been calculated as 69%, 41% and 29.3% respectively.

3.4 OVERALL PLANT CONFIGURATION FOR HYDROCARBON PRODUCTION (Cont'd)

For the ethylene production, the overall processing scheme is similar to that presented in Drawing 3-1 up to the acetylene hydrogenation unit. The exiting stream is then fed directly into the ethylene and hydrogen recovery unit shown in Figure 3-3. The overall material and utilities balance summary is given in Appendix B. The overall and specific product (ethylene) efficiencies are 72% and 42.7% respectively.

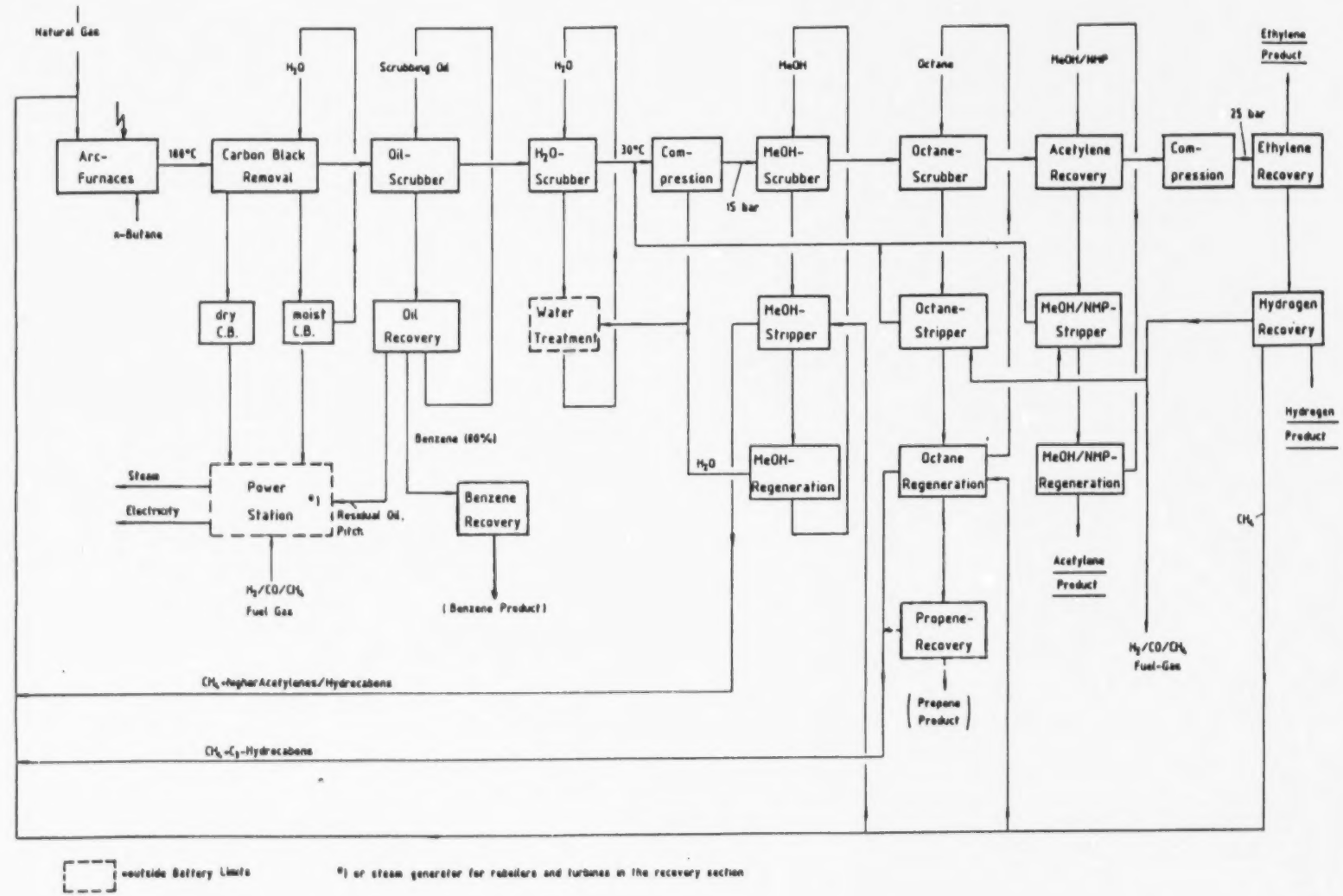
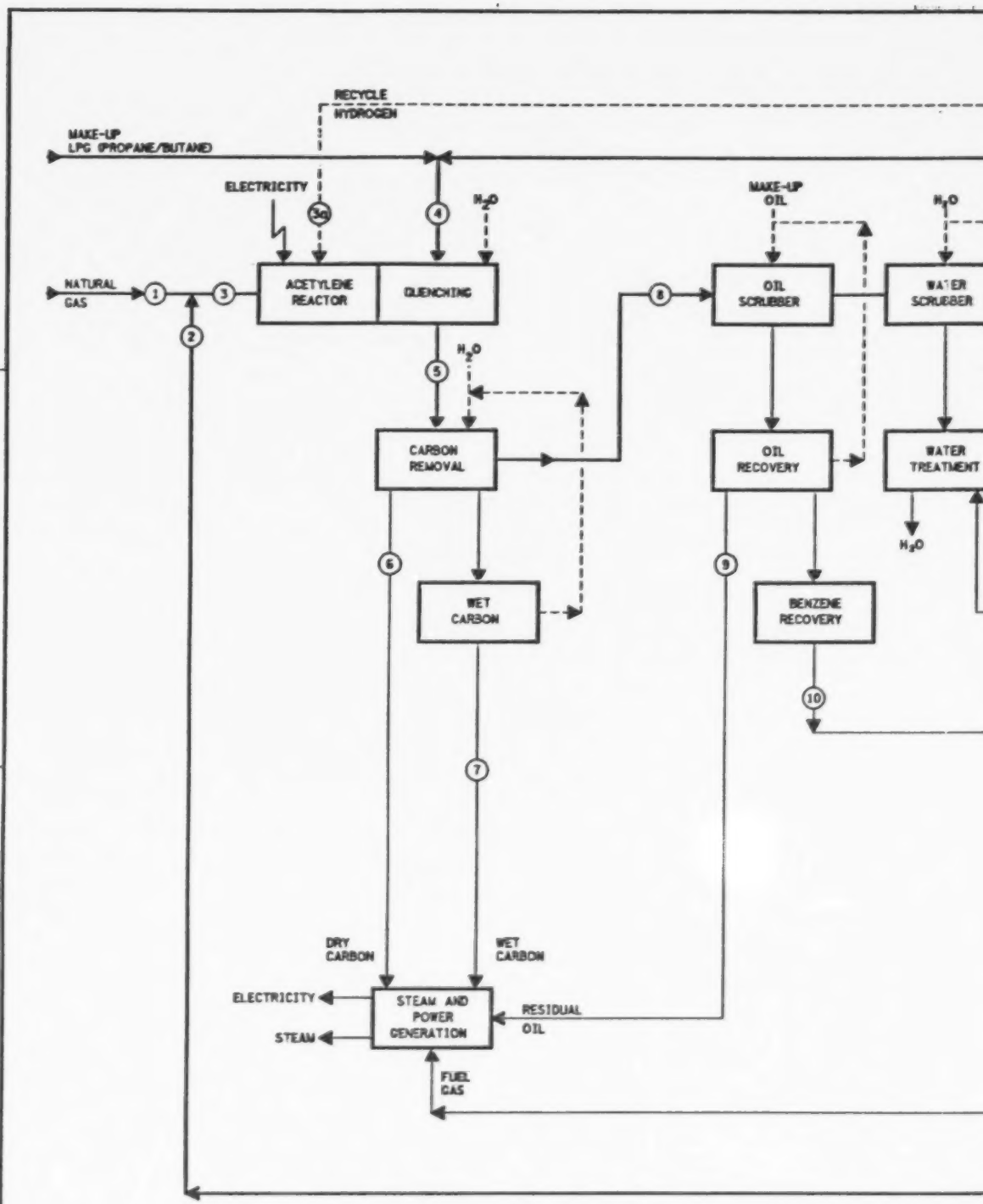
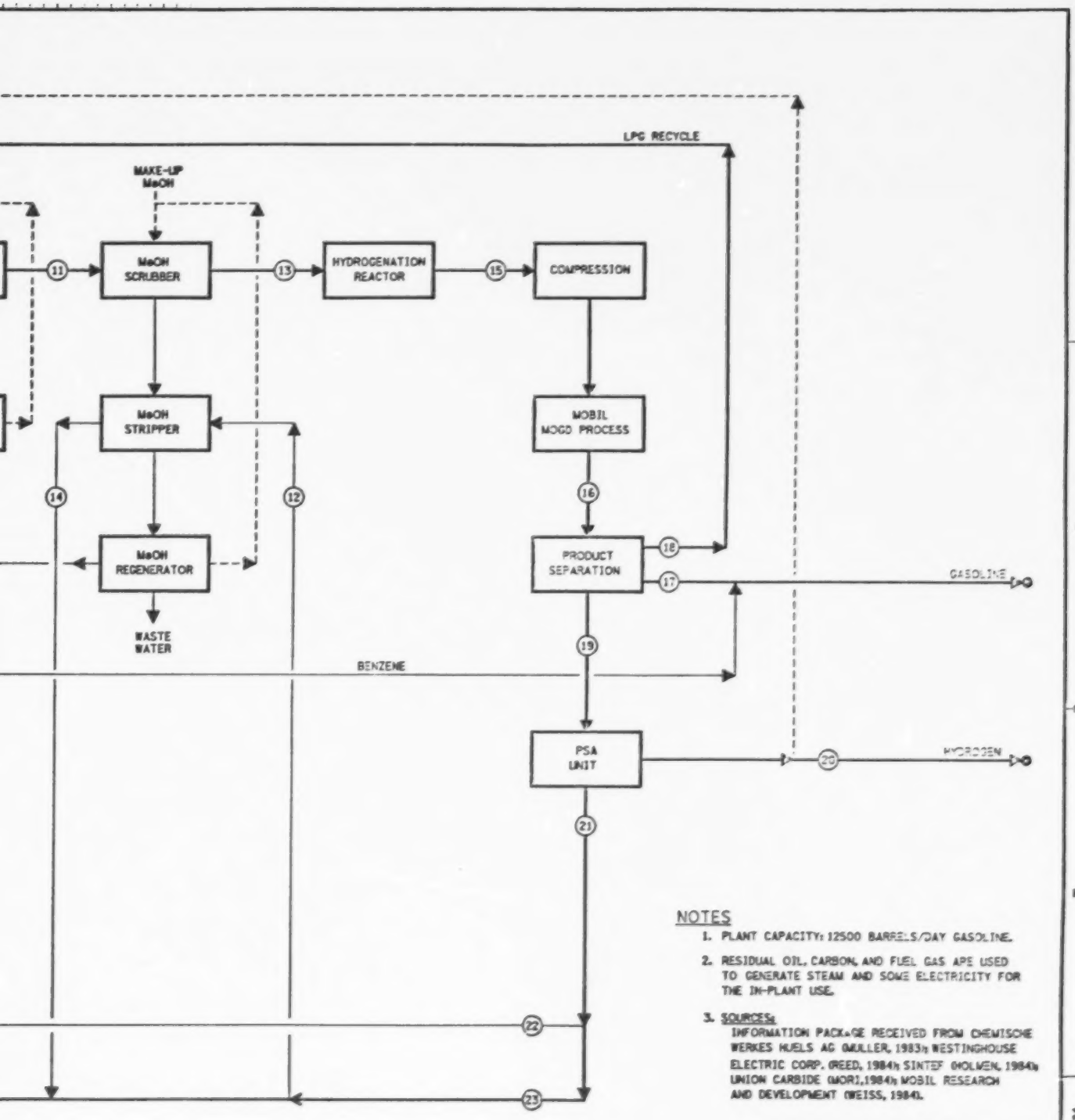


Figure 3-4: Simplified Flow Diagram for Acetylene Production Using the Arc-Acetylene Process (Chemische Werke Huels)



STREAM No	1	2	3	3a	4	5	6	7	8	9	10	11
STREAM DESCRIPTION FLOWRATE Kg/h	NATURAL GAS	RECYCLE GAS	TOTAL FEED	HYDROGEN	LPG QUENCH	REACTOR EFFLUENT	DRY CARBON	WET CARBON	GAS TO PRODUCT RECOVERY	RESIDUAL OIL	BENZENE	GAS TO MeOH SCRUBBER
ELECTRIC ARC REACTOR	81170	36165	117335	---	33865	151199	11246	5622	134332	4030	2061	129780
HYDROGEN PLASMA REACTOR	74483	29441	103845	7977	32192	144014	4936	2843	136234	3933	2371	131412
RESISTANCE-HEATED REACTOR	73042	23623	96665	---	29517	125183	6717	3869	117807	1091	751	117317



NOTES

1. PLANT CAPACITY: 12500 BARRELS/DAY GASOLINE.
2. RESIDUAL OIL, CARBON, AND FUEL GAS ARE USED TO GENERATE STEAM AND SOME ELECTRICITY FOR THE IN-PLANT USE.
3. SOURCES:
   INFORMATION PACKAGE RECEIVED FROM CHEMISCHE WERKES HUELS AG (MULLER, 1983); WESTINGHOUSE ELECTRIC CORP. (REED, 1984); SINTEF (HOLMEN, 1984); UNION CARBIDE (MORI, 1984); MOBIL RESEARCH AND DEVELOPMENT (WEISS, 1984).

12	13	14	15	16	17	18	19	20	21	22	23
STRIPPING GAS	GAS TO HYDROGEN-ATION UNIT	RECYCLE GAS TO REACTOR	OLEFIN RICH FEED	MOBIL REACTOR	GASOLINE	LPG RECYCLE	OFF GAS	HYDROGEN BY-PRODUCT	RECYCLE GAS	FUEL GAS	RECYCLE GAS TO REACTOR
7469	123003	15180	123003	123003	60278	19640	43085	10816	32669	2259	30010
7142	125127	12388	125127	125127	60015	19580	45532	7850	29705	5509	24196
5885	111822	10662	111822	111822	61388	15603	34831	10034	24796	5951	18846

SNC Inc.
1 Complexe Desjardins, Montréal, Canada

CLIENT: EMR, CANMET

PROJET: GASOLINE AND HYDROCARBON PRODUCTION FROM NATURAL GAS VIA ACETYLENE USING ELECTRICITY-BASED REACTORS

TITRE: OVERALL BLOCK FLOW DIAGRAM OF A CONCEPTUAL, COMMERCIAL PLANT (12500 bbl/d GASOLINE PRODUCT)

CONTRAT	CODE DU DESSIN	SÉRIE	RÉV
	DFMING	3.1	

CAD021170024

FORMAT A1

4.0 COST CONSIDERATIONS

Table 4-1 Basic Financial Assumptions and Economic Criteria

Plant capacity	: 12 500 bbl/d (1 987 350 litre/d)
Plant location	: Québec
Constant canadian dollars	: mid-1984 basis
Construction period	: 3 years
Capital investment schedule	

Year	%
1	22
2	48
3	30

Project and estimating contingency	: 10%
Financing	: 100% equity
Cost of capital	: 5% after taxes or about 9% before taxes
Tax rate	: 45%
Investment tax credit	: neglected
Plant operating eficiency	

Year	%
1	90
2-20	100

Plant life	: 20 years
Depreciation method	: straight line
Salvage value	: zero
Plant service factor	: 8000 hours/yr (=0.91)
Natural gas cost	: $ 3.86/GJ ($4.0/$10^6$Btu)
LPG (propane/butane)	: $166/$m^3$ (16.6¢/litre)
Electricity cost	: 1¢/kWh
By-product credit for H_2	: $4.74 - $7.58/GJ ($5-8/10^6Btu)
Chemicals used	
- Heavy Oil	: $189/t ($30/bbl)
- Methanol	: $268.5/t

4.0 COST CONSIDERATIONS

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION

Preliminary or budget-type capital cost estimate, operating, maintenance and product costs are presented for the designated conceptual, commercial plants producing 12 500 bbl/d (1 987 350 litres/d) of gasoline product. These costs have been developed in accordance with the techno-economic criteria outlined in Chapter 1 and based on the conceptual plant configurations described in Chapter 3.

4.1.1 Capital Cost Estimate and Total Investment

Equipment cost and investment data received from the manufacturers, process developers and licensors were taken as a basis for estimating the capital cost of the conceptual, commercial plants. Where information and data were not available, SNC in-house data or factored cost estimates were used. Analysis of the cost estimates obtained by SNC from various sources revealed that in many cases the philosophy and methodology, as well as the criteria applied, differed significantly. Therefore, for consistent techno-economic evaluation, these estimates were converted and adjusted for a Quebec location following the procedure described below and the criteria outlined earlier in Chapter 1. These criteria are reproduced as Table 4-1. All costs are calculated in "constant" mid-1984 Canadian dollars. Forward escalation beyond mid-1984 is not included.

Machinery and equipment are assumed to consist of 45% Canadian content. Items purchased abroad are assumed to be unavailable in Canada and are therefore exempted from customs duties. Costs and quotations obtained in U.S. dollars have been converted to

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.1 Capital Cost Estimate and Total Investment (Cont'd)

Canadian dollars using an exchange rate of 1.28 (Royal Bank of Canada, Foreign Exchange Quotations, May/June 84). Similarly, a currency exchange factor of 0.4562 was used for the West German Deutsche Mark.

Most of the bulk materials such as piping, fittings, concrete, etc., are assumed to be 100% Canadian. The U.S. quotes (Gulf Coast location) were converted to Canadian conditions (Quebec) using a material-price differential factor of 1.26. This factor was obtained from the standard cost indices and SNC in-house data, details of the calculations are given in Appendix C-1.

Construction manpower and labor estimates supplied by process developer for U.S. Gulf Coast locations were converted to a Quebec location using a productivity adjustment factor of 1.10 (see Appendix C-2). To convert from West German conditions, an adjustment factor of 1.14 was applied (Chauvel et al, IFP, Manual of Economic Analysis of Chemical Processes).

Machinery and equipment costs obtained from U.S. were adjusted for inflation to the base year (mid-1984) using Marshall and Stevens Equipment Cost index. For converting cost data of the complete process unit, Chemical Engineering Plant Index was used. For updating costs obtained from Canadian sources, Chemical and Petroleum Plant Price Index published by Statistics Canada was used. These indices are presented in Appendix C-3. In some cases, adjustment for equipment capacity was also made. Exponential factors used are given in Appendix C-4.

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.1 Capital Cost Estimate and Total Investment (Cont'd)

A winterization factor of 2% was applied to the total plant cost to reflect conditions in Québec which necessitate extra insulation, heat tracing in the pipes, etc.

The capital cost estimates for 12 500 bbl/d liquid fuel plant for each of the reactor configurations are presented in Table 4-2. For the electric arc reactor configuration, capital cost estimates supplied by Chemische Werke Huels (Germany) have been converted to Canadian conditions following the above procedure. Similarly, for hydrogen plasma and resistance-heated reactors, capital cost estimates are based on information supplied by Westinghouse Electric Corporation (USA) and SINTEF (Norway) respectively. Data for the MOGD process and PSA unit were supplied by Mobil Oil and Union Carbide respectively. The capital costs for offsite facilities have been calculated using SNC in-house data.

The total investment costs of the commercial liquid fuel plants are given in Table 4-3. The installed-plant cost includes all the equipment items for onsite and offsite facilities, bulk materials, labor, engineering and support, and other construction costs. A project and estimation contingency of 10% has been added to provide for an allowance in cost estimation, equipment damage during construction, delay in equipment delivery and project schedule.

To take account of other start-up and miscellaneous expenses during commissioning and the start-up period, an additional

Table 4-2 Capital Cost Estimate of 12 500 bbl/d Gasoline Plants *

Process Areas/Units	Millions of Canadian Dollars (mid-1984) Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Acetylene Production, Carbon Black Recovery and Higher Acetylene Separation	187.13	187.03	169.58
Acetylene Hydrogenation Unit	31.43	32.11	29.68
Mobil MOGD Process	68.11	68.93	63.71
PSA Unit for H_2 Recovery	25.14	26.05	21.47
Total Onsite Cost	312.11	314.13	284.44
Steam Generation	3.90	3.87	3.67
Incineration	2.95	2.21	2.07
Water Treatment	0.98	0.83	0.91
Electrical Equipment	89.98	106.00	90.73
Steam Cooling and Process H_2O	16.86	15.13	15.80
Refrigeration	8.04	6.71	7.40
Site Preparation, Building, etc.	21.57	21.75	20.14
Sewer, Fire Protection, Fence, Flare	4.61	4.65	4.30
Air Supply	1.60	1.41	1.33
Product Shipping and Storage	6.24	6.28	5.69
Inter Connecting Pipe Racks	6.53	7.03	6.33
Total Plant Cost	475.37	490.00	442.81

* The capital cost estimate includes 15% for engineering.

Sources:

Muller R., Chemische Werke Huels, Report to SNC Inc., Ref No. 1050/08/Bu, December 23, 1983

Reed, W.H., Westinghouse Electric Corp., Telex to SNC Inc., May 10, 1984

Holmen, A., SINTEF, Norway, Ref 21/AH/bde, January 20, 1984

Weiss, G.J., Mobil Research and Development Corp., Ref No. T-MOGD-STD, 597-00, July 30, 1984

Mori, B.H., Union Carbide Ltd., Telex to SNC Inc., September 24, 1984

Chauvel, et al., A.H., Institut Français du Petrole, "Manual of Economic Analysis of Chemical Processes", 1981.

Table 4-3 Total Capital Investment of 12 500 bbl/d Gasoline Plants

Components	Millions of Canadian Dollars (mid-1984) Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Direct Capital Investment			
Installed Plant Cost (TPC)	475.37	490.00	442.81
Project/Estimation Contingency	47.54	49.00	44.28
Total Installed Plant Cost (TIPC)	522.91	539.00	487.09
Indirect Capital Investment			
License and Prepaid Royalties	1.37	1.50	1.50
Initial Catalysts & Chemicals	1.35	1.35	1.31
Start-Up and Other Costs	15.69	16.17	14.61
Sub-Total	541.32	558.02	504.51
Working Capital*	15.69	16.17	14.61
Land*	1.09	1.09	1.09
Total Capital Investment (TCI)	558.10	575.28	520.21

*non-depreciable item

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.1 Capital Cost Estimate and Total Investment (Cont'd)

3% of the total installed plant cost (TIPC) has been included to allow for additional capital needed for any minor modifications required and process and environmental studies.

Cost of "initial charge" of catalysts and chemicals required for the process have been capitalized. Where data were not available, the cost of a 30-day supply based on the plant's operation was assumed.

Working capital has been taken as 3% of TPIC. This factor includes an allowance for 30-day operating supplies, chemicals, and other consumables (except water), spare parts and warehouse stock, a 30-day payroll burden and overhead expenses, and a 10-day storage of gasoline product.

Although the study is not site specific, land costs have been considered for a plot about 40 hectares (100 acres) at an average rate of $27 250/hectare ($2.70/m^2).

Interest during construction is not added to the plant investment, since for calculation purposes it was assumed that the capital required for the project will be provided through 100% equity financing. The total investment does not include the following:

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.1 Capital Cost Estimate and Total Investment (Cont'd)

- Process development cost and possible changes in design due to advancement of technology after commencement of contract
- Gas pipeline and distribution system outside the plant site
- Electricity generation and distribution outside the plant site
- Product distribution pipeline beyond battery limits
- Water supply and pipelines outside the plant site
- Housing, community facilities and associated infrastructure
- Roads and other transportation facilities.

4.1.2 Operating, Maintenance and Production Costs

Operating, maintenance and production costs are presented in Table 4-4. These costs are reported on an annual basis assuming plant operation of 8000 hours/year, i.e., a service factor of 0.91.

Raw material, utilities, catalysts and chemical costs have been calculated using unit cost data given in Table 4-1 and the annual estimated requirement presented earlier in Tables 3-1, 3-4 and 3.6. Make-up water is purchased at a cost of $\$0.132/m^3$ ($0.50/1000 US gallons).

Table 4-4 Annual Operating, Maintenance and Production Costs

Plant Capacity: 12 500 bbl/d Gasoline Product

Components	Millions of Canadian Dollars (mid-1984) Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Natural Gas	131.63	120.66	118.45
Electricity	42.56	55.20	43.84
LPG	34.71	30.77	37.30
Catalysts and Chemicals			
Scrubbing Oil	1.41	1.43	1.24
Methanol	0.99	1.01	0.90
Hydrogenation Catalyst	1.55	1.55	1.36
Mobil Catalyst*	12.25	12.25	12.25
Make-up Water	0.96	0.74	0.86
Running Royalties			
Acetylene Unit	2.04	2.24	2.24
Mobil MOGD Process	*	*	*
Labor and Supervision	1.97	1.97	2.08
Operating Supplies	0.30	0.30	0.37
Maintenance Cost	15.69	16.17	14.61
Property Taxes and Insurance	11.76	12.13	10.96
Overhead and Administration	6.18	6.35	5.84
Total Operating Cost	263.95	262.77	252.30
By-product Credits for H_2 at $7.58/GJ	(77.89)	(56.53)	(72.26)
Net Operating Cost	186.06	206.24	180.04
Depreciation	27.07	27.90	25.23
Capital Recovery (ROE, taxes,etc)	35.19	39.57	33.29
Total Net-Production Cost (By-product Credits at $7.58/GJ)	248.32	273.71	238.56

Table 4-4 Annual Operating, Maintenance and Production Costs (Cont'd)

Plant Capacity: 12 500 bbl/d Gasoline Product

	Millions of Canadian Dollars (mid-1984)		
Components	Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Total Operating Cost	263.95	262.77	252.30
By-product Credits for H_2 at $4.74/GJ	(48.68)	(35.33)	(45.16)
Net Operating Cost	215.27	227.44	207.14
Depreciation	27.07	27.90	25.23
Capital Recovery (ROE, taxes, etc)	35.19	39.57	33.29
Total Net-Production Cost (By-product credits at $4.74/GJ)	277.53	294.91	265.66

* For proprietary reasons, running royalties are included in the annual catalyst cost.

Sources: Muller R., Chemische Werke Huels, Report to SNC Inc., Ref No. 1050/08/Bu, December 23, 1983

Reed, W.H., Westinghouse Electric Corp., Telex to SNC Inc., May 10, 1984

Holmen, A., SINTEF, Norway, Rej 21/AH/bde, January 20, 1984

Weiss, G.J., Mobil Research and Development Corp., Ref No. T-MOGD-STD, 597-00, July 30, 1984

Mori, B.H., Union Carbide Ltd., Telex to SNC Inc., September 24, 1984

Chauvel, et al., A.H., Institut Français du Petrole, "Manual of Economic Analysis of Chemical Process", 1981.

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.2 Operating, Maintenance and Production Costs (Cont'd)

Running royalties have been estimated using information supplied by process developers and licensors. For acetylene production, running royalties are considered for 10 years and none thereafter. For proprietary reasons, running royalties for the Mobil MOGD process have been included in the annual cost of the catalyst.

Operating and supervisory labor costs are calculated from manpower requirements specified by the respective process developers. Where complete information was not available, typical manpower distribution given in Appendix C-5 was used. For annual disbursements, average hourly rates were multiplied by a factor of 1940 hours/year.

	Electric Arc Reactor	Hydrogen Plasma Reactor	Resistance Heated Reactor
• Operating Labor ($14/h)			
Onsite	26	26	27
Offsite	11	11	11
	37	37	38
• Supervision ($16.8/h)			
Onsite	9	9	10
Offsite	5	5	6
	14	14	16

A labor burden of 35% has been added to the direct operating and supervisory labor costs to cover fringe benefits, unemployment insurance, retirement plan, shift differentials, etc.

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.2 Operating, Maintenance and Production Costs (Cont'd)

Operating supplies are estimated at 30% of direct operating labor costs. The annual maintenance cost was taken as 3% of the total installed plant cost (TIPC). This includes maintenance materials and supplies, equipment replacement cost, maintenance labor, supervision, labor burden and other regular contract maintenance.

Property taxes and insurance costs are assessed at 2.25% of TIPC, comprising 1.5% for taxes and 0.75% for insurance.

Plant overhead and administration is assumed to be 35% of direct labour cost, operating supplies and maintenance. It includes plant management, engineering, fire protection, medical services and personnel, accounting, clerical and janitorial services.

To obtain net operating costs, by-product credits have been taken for hydrogen at two values: \$7.58/GJ (\$8.00/10^6 Btu) and \$4.70/GJ (\$5/10^6 Btu).

Levelized capital and investment related charges (depreciation, return on equity, taxes, etc.) are added to the net operating and maintenance costs to obtain the total production cost. These costs actually vary over the life of the project, however, equal annualized or levelized costs have been calculated using the present value method.

A straight line depreciation is applied over a 20-year plant life. A capital recovery component has been calculated for 100% equity financing assuming a 5% capital cost after tax (or about

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.2 Operating, Maintenance and Production Costs (Cont'd)

9% before tax) in "real" terms* for the "constant" dollar financial analysis. The levelized capital cost factor also takes into account the following:

- start-up efficiency of 90% in the first year of production and 100% in subsequent years
- running royalties for the acetylene unit for the first ten years and none thereafter
- zero salvage value of the depreciable component of the total plant investment
- terminal worth and recovery of non-depreciable components (land and working capital) at the end of the plant life.

The total net-production costs are shown for two cases, i.e., when product credits for hydrogen are taken at \$7.58/GJ (\$8.00/10^6 Btu) and \$4.74/GJ (\$5/10^6 Btu) respectively.

4.1.3 Product Cost

The unit product cost or (transfer price) of gasoline for each of the designated cases is shown in Table 4-5. A breakdown of product cost into various cost components is presented.

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* for "constant" analysis, cost of capital is taken in "real" terms, i.e. current rate minus inflation (14% - 5% = 9% before tax)

Table 4-5 Product Cost of Gasoline
Plant Capacity: 12 500 bbl/d gasoline product

	Millions of Canadian Dollars (mid-1984)		
	Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
At $7.58/GJ By-product Credits for H_2			
Product Cost (Transfer Price), ¢/litre			
Natural Gas & LPG	25.1	22.8	23.5
Electricity	6.4	8.3	6.6
Catalysts, Chemicals, Make-up H_2O & Royalties	2.9	2.9	2.8
Labor, Supervision, & Op. Supplies	0.4	0.4	0.4
Maintenance	2 4	2.4	2.2
Property, Taxes, Insurance & Overheads	2.7	2.8	2.6
	39.9	39.6	38.1
Credits for Hydrogen	(11.8)	(8.5)	(10.9)
Depreciation	4.1	4.2	3.8
	32.2	35.3	31.0
Capital Recovery Cost (ROE, taxes, etc.)	5.3	6.0	5.0
	37.5	41.3	36.0
At $4.74/GJ By-product Credits for H_2			
Product Cost (Transfer Price), ¢/litre			
	39.9	39.6	38.1
Credits for Hydrogen	(7.4)	(5.3)	(6.8)
Depreciation	4.1	4.2	3.8
	36.6	38.5	35.1
Capital Recovery Cost (ROE, taxes, etc.)	5.3	6.0	5.0
	41.9	44.5	40.1

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.3 Product Cost (Cont'd)

The major cost element is for raw materials and process input such as natural gas, LPG and electricity which account for about 80% of the net product cost. Other significant components are by-product credits for hydrogen, depreciation and capital recovery cost. If by-product credits are taken at a lower value, then product cost increases marginally by about 4¢/litre.

4.1.4 Techno-Economic Comparison of the Processes

A summary of techno-economic results is presented in Table 4-6. It may be seen that the conversion of natural gas is in the range of 80-84%, being highest for the resistance-heated reactor. Similar trends are observed for methane conversion, reactor product yield and SER. However, the overall process SER is nearly the same for the electric-arc and resistance-heated reactors. Similarly, the overall process thermal efficiency is also nearly the same, though the efficiency of the "preheated-hydrogen" reactor is somewhat lower. On the other hand, the electric-arc reactor has been commercialized for many years whereas the resistance-heated reactor has been tested only on the bench-scale and pilot-plant levels.

Total capital investment is nearly the same (within the accuracy of the cost estimate) for all three reactor configurations. However, the proposed novel route is less capital intensive as compared with other synfuel technologies, e.g. the Fisher-Tropsch process or direct coal liquefaction route. Comparisons with other direct coal liquefaction processes show that the investment

Table 4-6 Techno-Economic Comparison of the Designated Processes
Plant Capacity: 12 500 bbl/d gasoline product

	Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Natural Gas Conversion, wt%	79.5	79.5	83.0
CH_4 Conversion[1], Prequench, wt%	82.0	82.0	88.0
After quench*, wt%	74.0	74.0	80.0
Reactor Product Yield (Pre LPG Quench), wt%			
. Acetylene Yield[1]	58.5	66.6	79.7
. Acetylene + Ethylene Yield[1]	68.0	77.5	83.7
Reactor SER (Pre LPG Quench) **			
. kWh/kg of acetylene	11.3	14.5	9.9
. kWh/kg of acetylene + ethylene	9.7	12.5	9.4
Reactor SER (After LPG Quench)			
. kWh/kg of acetylene	11.3	14.5	9.9
. kWh/kg of acetylene + ethylene	8.0	10.3	8.0
Process Input, kg/kg of gasoline			
. Natural Gas	1.30	1.19	1.17
. LPG	0.23	0.20	0.25
Overall Process SER, kWh/kg of gasoline	10.0	12.5	10.2
By-Products Output, kg/kg of gasoline			
. Hydrogen	0.17	0.13	0.14
Total Energy Input, GJ/h	6 909	7 040	6 594
Total Energy Output, GJ/h	4 169	3 819	4 067
Thermal Efficiency, %			
. Overall	60.3	54.2	61.7
. Liquid	41.8	41.0	43.6

* Lower overall CH_4 conversion (after quench) is due to some methane formation during LPG quenching.

** Reactor SER (after LPG quench) based on acetylene is the same as SER (pre LPG quench) since negligible acetylene is formed during quenching.

Table 4-6 Techno-Economic Comparison of the Designated Processes
Plant Capacity: 12 500 bbl/d gasoline product (Cont'd)

	Using Electric-Arc Reactor	Using Hydrogen Plasma Reactor	Using Resistance Heater Reactor
Development Status	Commercial	Pilot Plant	Bench-Scale Pilot Plant
Capital Investment, $ million			
. Direct	522.91	539.00	487.09
. Indirect	35.19	36.28	33.12
Total	558.10	575.28	520.21
Investment, $/daily litre of gasoline	280.8	289.5	261.8
Gasoline Product Cost, ¢/litre			
. At $7.58/GJ H_2	37.5	41.3	36.0
. At $4.74/GJ H_2	41.9	44.5	40.1
Breakdown of Product Cost, %			
. Process Input[2]	84.0	75.3	83.6
. Operating Cost[3]	15.2	13.2	15.0
. Overheads[4]	7.2	6.8	7.2
	106.4	95.6	105.8
. By-products Credits	(31.4)	(20.6)	(30.3)
	75.0	75.3	75.5
. Depreciation	10.9	10.2	10.6
. Capital Recovery Cost	14.1	14.5	13.9
	100.0	100.0	100.0

Notes: 1. Based on total CH_4 feed to the reactor (i.e. including CH_4 in the recycle gas stream).

2. Process input includes natural gas, LPG and electricity.

3. Operating cost includes catalyst and chemicals, make-up water, running royalties, labor, supervision, operating supplies and maintenance cost.

4. Overheads include property taxes, insurance, general overhead and administration expenses.

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.4 Techno-Economic Comparison of the Processes (Cont'd)

required is less than 50%, as shown below:

	Proposed Route	Direct-Coal Liquefaction
• Investment, $/daily litre of product	260 - 280	500 - 800

Net product costs range from 36 to 41.3¢/litre, being nearly the same for the electric arc and resistance-heater reactor configurations. Analysis of the product costs show that raw materials and process input represent about 80% of the final gasoline cost. Therefore, improvement in gas conversion, lowering of reactor SER, better heat recovery and lower cost of raw materials are factors which can have significant impact on the net product cost. For example, natural gas has been taken at full border price ($3.86/GJ). A 25% lower cost than the border price will reduce the gasoline cost by 5¢/litre. A 20% higher cost of electricity (1.2¢/kWh) will increase the gasoline cost by 1.1 to 1.8¢/litre depending on the reactor configuration.

Comparing various reactor configurations, the so-called "isothermal" resistance-type reactor has the obvious advantage of easier temperature control and more uniform temperature profile in the reactor, thus making it possible to obtain high yields of acetylene and a higher conversion of natural gas. Also, the electrical energy is converted into thermal energy at the main frequency and voltage without the complex and relatively expensive rectification, etc., needed in the electric-arc or plasma reactor configurations.

4.1 CAPITAL, OPERATING AND PRODUCT COSTS FOR LIQUID FUEL PRODUCTION (Cont'd)

4.1.4 Techno-Economic Comparison of the Processes (Cont'd)

On the other hand, for continuous commercial-scale application for acetylene production from hydrocarbon, no satisfactory resistor is known; they are either embrittled by the high temperatures, (Pt and alloy) or form carbides (Mo, T, W) or the material (SiC) is damaged and embrittled by hydrogen. A carbon resistor will operate satisfactorily but will "grow" as the formation of carbon during acetylene production is autocatalytic. Deposition of carbon will also lower the efficiency of heat transfer to the feed gas and cause reactor blockage problems. For prolonged steady operation, the resistor may be isolated from the reacting gas by use of a suitable refractory lining. This will necessitate the use of a higher temperature in the surrounding resistor to achieve a large heat flux, thus requiring more energy and higher SER. Also, the gas passage which will have to be narrow may get easily choked by a small amount of carbon deposition if operating parameters are not closely controlled. Normal carbon deposits may be eliminated by periodic combustion with air.

The resistance-type reactor has so far been tested only on the bench-scale and pilot-plant levels. Further process development using this reactor configuration has been abandoned due to financial constraints [Holmen 1984]. Therefore, the commercial potential of this reactor in the near future is low.

The electric-arc reactor configuration of Chemische Werke Huels has been extensively tested and commercialized for over 45 years. Operational difficulties associated with the stability

4.1.4 **Techno-Economic Comparison of the Processes** (Cont'd)

of the arc and carbon deposits have been resolved. The electric arc process at Huels is commercially proven; therefore no technical risk is involved.

The hydrogen plasma reactor, though not extensively listed, is similar in concept to the electric-arc reactor; however, it offers some operational advantages with reduced carbon formation and less maintenance requirements.

4.2 PRODUCT COSTS OF DESIGNATED HYDROCARBONS

For many years, acetylene has been used as a major building block for the production of large-volume commodity chemicals as well as lower-volume Reppe acetylene-based products and specialty acetylenic alcohols, diols and other derivatives. In the early 1960's, the use of acetylene for the production of commodity chemicals (acrylonitrile, acrylates, chlorinated solvents, chloroprene, vinyl acetate, vinyl chloride) reached a peak of about 1 billion pounds of acetylene [McGreevey 1975]. Since then, acetylene has been largely displaced by olefins such as ethylene, propylene, and butadiene derived from cheap cracking feedstocks. These olefins, though less suitable as feedstocks than acetylene for some applications, were still used as they became readily available at prices considerably lower than acetylene. Ethylene is the main competitor of acetylene.

On the other hand, the use of acetylene for industrial purposes such as welding, metal cutting and fabrication has remained constant. However, there has been a significant growth in Reppe-type chemicals such as butyndiol, butanediol, tetrahydrofuran (THF), n-methyl-2-pyrrolidone, vinylpyrrolidone, and polyvinyl pyrrolidone, besides a variety of smaller volume acetylenic alcohols and glycols and their derivatives [Ring 1978].

As a sub-objective, the feasibility of hydrocarbon production was investigated for these designated cases:

- acetylene production as a feedstock for
  - Reppe-type chemicals, 50 million lb/yr (22.7 Gg/yr) acetylene

4.2 PRODUCT COSTS OF DESIGNATED HYDROCARBONS (Cont'd)

- Commodity chemicals, 300 million lb/yr (136.2 Gg/yr acetylene)

. ethylene as a feedstock for petrochemical production, 1 billion lb/yr (454 Gg/yr) ethylene

Preliminary capital, operating and product costs are summarized in Table 4-7. These estimates are based on the techno-economic data supplied by Chemische Werke Huels [Muller 1983] for the plant units within the plant battery limits and those developed by SNC for the associated offsite facilities. Similar techno-economic criteria and procedures described earlier for the liquid fuel production (see Table 4-1) have been applied for the acetylene and ethylene plants except that for a smaller acetylene plant (50 million lb/yr), electricity rates are taken at 1.2¢/kWh. By-product credits are taken for ethylene at 25¢/lb (55¢/kg),* hydrogen at $5/10^6 Btu ($4.74/GJ), propylene and butylene at 19¢/lb (42¢/kg), benzene at $1.80/imp. gal (53.9¢/kg).

Operating and supervisory labor costs are calculated from the manpower requirements specified by the process developers. For annual disbursements, average hourly rates were multiplied by a factor of 1940 hours/year.

	Acetylene Plant		Ethylene Plant
. Operating Labor ($14/h)			
Onsite	8	18	35
Offsite	4	8	14
	12	26	49

* for acetylene plant only

Table 4-7 Capital, Operating and Product Costs for Hydrocarbon Production
Plant Location: Quebec
Reactor Configuration: Electric-Arc (Chemische Werke Huels)

	Millions of Canadian Dollars (mid-1984)		
	Acetylene Plant		Ethylene Plant
Plant Capacity million lb/yr	50	300	1000
Direct Capital Investment			
Installed Plant Cost (TPC)	67.80	239.76	456.20
Project/Estimation Contingency	6.78	23.98	45.62
Total Installed Plant Cost (TIPC)	74.58	263.74	501.82
Indirect Capital Investment			
Licence and Prepaid Royalties	1.37	1.37	2.70
Initial Catalysts & Chemicals	0.10	0.20	0.51
Start-Up and Other Costs	2.23	7.91	15.05
Sub-Total	78.28	273.22	520.08
Working Capital*	2.24	7.91	15.05
Land*	0.54	0.65	1.30
Total Capital Investment (TCI)	81.06	281.78	536.43
Annual Operating, Maintenance and Production Costs			
Raw Material and Process Input	16.04	92.8	214.72
Operating Cost	4.45	13.0	26.77
Overheads	2.68	9.15	17.39
	23.17	114.95	258.88
By-products Credits	(10.74)	(64.42)	(83.52)
	12.43	50.53	175.36
Depreciation	3.91	13.66	26.00
Capital Recovery Cost (ROE, taxes, etc.)	5.34	16.45	41.92
	21.68	80.94	243.28
Product Cost			
¢/lb	43.4	26.9	24.3
¢/kg	96.0	59.4	53.6

* non-depreciable items

4.2 PRODUCT COSTS OF DESIGNATED HYDROCARBONS (Cont'd)

	Acetylene Plant		Ethylene Plant
• Supervision ($16.8/h)			
Onsite	4	9	11
Offsite	2	5	7
	6	8	18

For a 50-million lb/yr plant, the acetylene product cost (transfer price) is about 43.4¢/lb compared to 27¢/lb for the larger plant. These costs make acetylene attractive as a feedstock for chemical production. Also, the results indicate that, as in the liquid fuel production, the raw materials account for the major portion of the final product cost. For this study, a full border price was assumed for the natural gas and LPG; a discount on raw material price would have a significant impact on the net product cost. For example, a 25% discount would reduce the product cost by about 4¢/lb.

5.0 CONCLUSIONS

5.0 CONCLUSIONS

A review and evaluation of the three designated acetylene reactors was carried out for production of liquid fuels (gasoline) from natural gas. A conceptual design, overall material and energy balance with utilities requirements was prepared for each process configuration. Budget-type capital, operating and product costs have been prepared.

Based on the analysis of data provided by process developers and the techno-economic evaluation the following conclusions are made:

- Rapid heating followed by rapid quenching is essential for desired conversion and acetylene yield.

- Direct quenching with a suitable quenchant is not only an efficient method of stabilizing the acetylene product but also serves to recover part of the process heat. Indirect heat recovery by use of pebble coolers or checker-brick furnaces does not appear to be a satisfactory alternative.

- The designated acetylene reactors are sensitive to process variations, however, stability of the arc can be maintained by either "rotating" the gas stream through a tangential feed chamber as in the electric-arc reactor or by rotating the arc itself as in the hydrogen-plasma reactor configuration. There is no inherent advantage of one method over the other, although recent developments favor the latter approach.

5.0 CONCLUSIONS (Cont'd)

- In spite of a non-uniform temperature profile in the arc zone, particularly the steep radial temperature gradient as in the electric-arc/hydrogen-plasma reactor, there is no intrinsic difficulty associated with maintaining the "average" conversion and composition of the "total" product gas stream.

- The resistance-heated reactor has an inherent advantage over the electric-arc/hydrogen plasma reactors. It is easier to control and maintain a uniform temperature profile, thus allowing higher conversion and acetylene product yield.

- Overall conversion of natural gas ranges from 80-83% depending on the reactor configuration chosen, being highest for the resistance-heated reactor. Methane conversion ranges from 82% for electric-arc/plasma reactor to 88% for resistance-heated type reactor.

- Higher acetylene yields are obtainable in the resistance-heated rector (80%) as compared to the hydrogen plasma reactor (67%) and the electric-arc reactor (58.5%).

- Similar trends are observed for the reactor specific energy requirements (9.9-14.5 kWh/kg of acetylene formed and 9.4-12.5 kWh/kg of acetylene and ethylene formed). However, overall process SER is nearly the same for resistance-type and electric-arc reactor (10 kWh/kg of gasoline), though for the hydrogen-plasma reactor the SER is slightly higher (12.5 kWh/kg of gasoline).

5.0 CONCLUSIONS (Cont'd)

- Similarly, the overall thermal efficiency of electric-arc and resistance-type reactor configurations is nearly the same (about 61%), though for hydrogen-plasma reactor it is somewhat lower (54%). The "liquid" product efficiency, however, is nearly the same for all three reactor configurations (42-43.6%).

- The resistance-type reactor has been tested only on bench-scale and pilot plant levels, therefore significant research and development may be required prior to commercialization. At the present level of development, there is a high degree of technical risk involved. Also, undesirable carbon formation in the narrow annular space may lead to significant operational problems.

- The electric-arc reactor is commercially proven. The hydrogen plasma reactor has also been extensively tested.

- The novel scheme for liquid fuel production via the acetylene route offers an attractive synfuel approach for liquid fuel and/or hydrocarbon production. The proposed method involves utilization of both surplus natural gas and electricity.

- The product cost (transfer price) of gasoline is nearly the same for all three reactor configurations ranging from 36-41.3¢/litre for the base case.

- The major cost components are raw material and process input, accounting for about 80% of the net product cost. Other significant components are by-product credits for hydrogen, depreciation and capital recovery cost.

5.0 CONCLUSIONS (Cont'd)

- Although the scope of the study did not include the sensitivity analysis, a 25% reduction in the cost of natural gas from the full border price of $3.86/GJ would reduce the gasoline product cost by 5¢/litre. A 20% increase in electricity cost (1.2¢/kWh) will result in 1.1 to 1.8¢/litre in gasoline product cost depending on the reactor configuration chosen. Other industry incentives, e.g. tax credits, would have further positive impact on the final product. These incentives were not considered at this stage of the study.

- Compared to existing cost of gasoline (32 - 35¢/litre) from conventional technology, the estimated cost of gasoline is attractive. It is recommended that further study be done to explore the potential of making gasoline via the novel acetylene route.

- The proposed route is less capital intensive. The capital investment of $260-280 per daily litre of product is less than 50%, compared to other synfuel technologies, such as Fischer-Tropsch or conventional coal liquefaction technologies.

- With the exception of the MOGD process, all the supporting process units have been extensively tested and commercially available. The MOGD process has been tested mainly for the production of liquid fuel and/or distillates from higher olefin (C_3^-) feed. For lighter olefin feed (C_2^-), as in our case, further testing and development may be required.

For the hydrocarbon option, the following conclusions are made:

5.0 CONCLUSIONS (Cont'd)

- Product cost (transfer price) of acetylene for a 50 million lb/yr capacity is estimated at 96¢/kg. Total investment required is $81 million. At this cost, acetylene production for smaller-volume speciality chemicals is very attractive. A possible discount of 10% on natural gas would further reduce the transfer price by 6¢/kg.

- For large-volume commodity chemicals, (300 million lb/yr acetylene capacity) the product cost of acetylene is about 60¢/kg. The total plant investment is $282 million. At this price, the acetylene price is competitive with ethylene.

- Production of ethylene itself via acetylene is not economically attractive. The product cost of ethylene for 1 billion lb/yr plant is calculated to be 54¢/kg which is significantly higher in comparison to the cost of ethylene obtained from a conventional ethylene process (37-45¢/kg) using natural gas as a feedstock.

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APPENDIX A

THERMODYNAMICS AND KINETICS CONSIDERATION FOR ACETYLENE PRODUCTION

APPENDIX A: THERMODYNAMICS AND KINETICS CONSIDERATION FOR ACETYLENE PRODUCTION

In general, the method of producing acetylene by thermal pyrolysis of hydrocarbon takes advantage of the fact that the free energy of acetylene decreases at higher temperatures in contrast to the behavior of other hydrocarbons. Figure A-1 is a plot of free energy of formation of various hydrocarbons including acetylene (C_2H_2) as a function of temperature.

At 1500 K or higher, acetylene has lower free energy (more stable) than any other hydrocarbon, therefore acetylene becomes a predominant specie at higher temperatures. But even at these high temperatures, acetylene is less stable than its free elements (carbon and hydrogen). The curves in Figure A-1 also indicate that the temperature for the formation of acetylene decreases with an increase in the molecular weight of the paraffinic hydrocarbons.

The amount of energy required to convert the hydrocarbons is very large and must be applied at high temperatures to achieve a favorable free energy relation. A calculation of the theoretical energy requirement has been reported by Miller [1966] based on the following general equation and a reaction temperature at which Kp for the reaction was 100:

$$\frac{2}{n} C_n H_{2n+2} \longrightarrow C_2H_2 + \left(1 + \frac{2}{n} H_2\right)$$

Hydrocarbons	Reaction Temp, C	Heat of Reaction at 25 C, kJ/mol*	Sensible Heat kJ/mol*	Heat Required per Unit of C_2H_2 kJ/mol*	kWh/kg
Methane	1470	376.6	227.6	604.2	6.45
Ethane	1160	311.3	144.3	455.6	4.86
Propane	1090	295.8	123.8	419.6	4.48
Butane	1000	289.1	114.2	403.3	4.31

* mol = gram mole

APPENDIX A: THERMODYNAMICS AND KINETICS CONSIDERATION FOR ACETYLENE PRODUCTION (Cont'd)

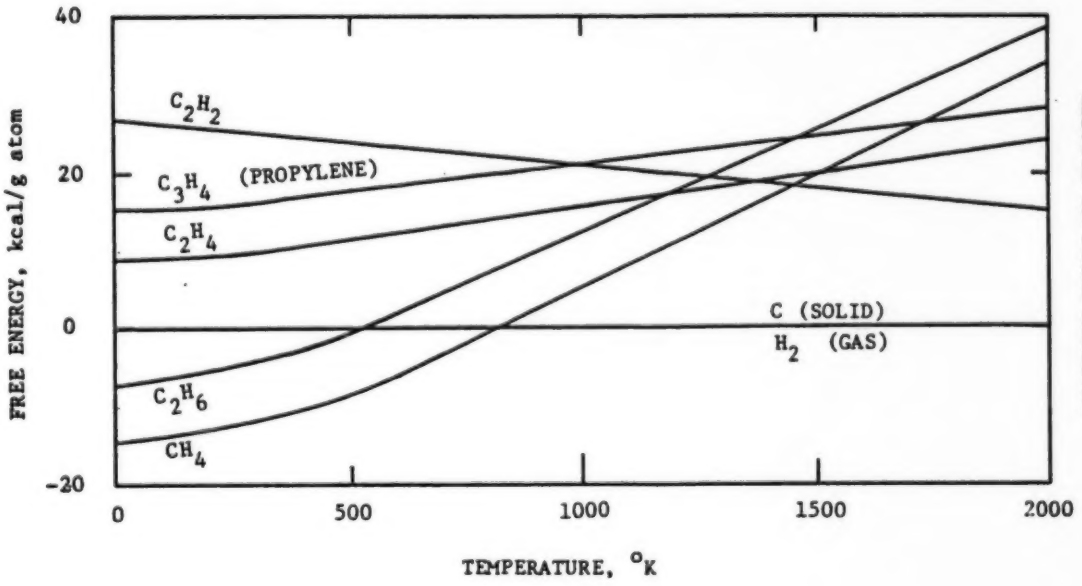
In addition, to the tabulated theoretical heat requirement, energy must also be supplied for:

- heat losses (in water-cooled electrodes, radiation to the reactor chamber walls)
- enthalpy of unreacted feedstock
- heat supplied to inert elements present in feedstock, e.g. N_2
- energy requirement for the side reaction, such as formation of olefins (e.g. ethylene) or of carbon (e.g. thermaldecomposition of $C_2H_2 \rightarrow 2C + H_2$ is exothermic), and
- energy requirement for plant operation

The thermodynamic data show that with an increasing chain length, the energy requirements less to convert hydrocarbons to acetylene decreases. Also, typical data for heat of formation of various hydrocarbons divided by the number of carbon atoms in the hydrocarbon support the above results as shown below:

	kJ/carbon atom at 1500 C
methane	-92.4
ethane	-53.9
n-octane	-31.9
carbon	0
ethylene	+18.1
acetylene	+110.1

The decomposition of hydrocarbons (mainly acetylene) to carbon and hydrogen begins at a relatively low temperature. In the case of methane, for example, this occurs at about 600 C [Kirk and Othmer 1964]. Hence, the decomposition proceeds in competition with the formation of acetylene. However, the rate of acetylene formation is greater than that of the decomposition reaction [Leroux 1961, Palmer 1964]. Therefore, the commercial manufacture of acetylene from pyrolysis of hydrocarbons must involve rapid heating of the feedstock, followed by rapid quenching of the product to minimize the decomposition of acetylene formed.



Note: 1 kcal = 4.184 kJ

Figure A-1: Free Energy of Formation of Some Hydrocarbons

APPENDIX - B

OVERALL MATERIAL AND UTILITIES BALANCE SUMMARY FOR DESIGNATED HYDROCARBON PRODUCTION

Appendix B Overall Material and Utilities Balance Summary for the Designated Hydrocarbon Production

	Acetylene Plant		Ethylene Plant
Plant Capacity million lb/yr	50	300	1000
Raw Material and Process Input, kg/h			
Natural Gas	4 616	27 695	63 845
LPG	2 050	12 297	28 346
Scrubbing Oil	210	425	784
Methanol	210	425	784
Octane	105	212	378
Electricity, MW			
Arc Furnace	32	194	447
Plant Use	5	29	67
	37	223	514
Utilities Required			
* HP Steam (2 MPa), kg/h	7 373	44 236	101 973
* LP Steam (0.4 MPa), kg/h	3 970	23 820	54 810
** Cooling Water (t_e = 26°C, Δt = 10°C)	1 700	10 208	23 532
Make-up Water, m^3/h	230	325	750
** Instrument Air (0.5 MPa, Dew point = -25°C), Nm^3/h	85	510	1 175
** Nitrogen (0.4 MPa), Nm^3/h	85	510	1 175

Appendix B Overall Material and Utilities Balance Summary for the Designated Hydrocarbon Production (Cont'd)

	Acetylene Plant		Ethylene Plant
Plant Capacity million lb/yr	50	300	1000
Products/By-products, kg/h			
Acetylene	2 836	17 014	-
Ethylene	1 118	6 706	58 175
Hydrogen	1 014	6 086	14 730
Propene/Butene	210	1 263	2 911
Benzene	125	748	1 725
* Residual Oil	244	1 463	3 374
* Carbon Black	1 021	6 125	9 672
* Fuel Gas	243	1 456	9 672
Thermal Efficiency, %			
Overall Product	69.0	69.0	72.0
Desired Product (C_2H_2 + C_2H_4)	40.9	40.9	--
Specific Product (C_2H_2 or C_2H_4)	29.3	29.3	42.7

* Residual oil, carbon black and fuel gas are used to generate steam and some electricity for in-plant use.

** Cooling water, instrument air and nitrogen are produced within the plant.

Source:
Muller, R., Chemische Werke Huels, Report to SNC Inc., Ref. No. 1050/08/Bu, December 23, 1983

APPENDIX - C

SOME ECONOMIC AND FINANCIAL ASSUMPTIONS AND DATA

Appendix C-1 Canada/USA Material Price Differential

Reference Location: Gulf Coast (USA)

Plant Location : Quebec (Canada)

	Relative Price*	Material** Distribution %	
Concrete Mix	1.37	30	0.411
Standard Structure			
Wide Flanges	1.27	7.5	0.095
I-Beams	1.21	7.5	0.095
Piping			
Iron	1.2	36.0	0.432
Copper	1.3	9.0	0.117
Cables/Wires	1.10	10.0	0.11
		100.0	1.26

Sources:

* Engineering News Record, June 4, 1984; July 5, 1984

** Based on SNC in-house data on various petrochemical projects.

Appendix C-2 Productivity Adjustment Factor

A review of labor productivity based on petrochemical projects indicates that a multiple of 1.05 be applied for conversion of U.S. Gulf Coast labor to equivalent Canadian labor.*

For adjustment to the Quebec region, climatic factors must also be considered. Therefore, an allowance of 12 days per year has been provided for adverse working conditions during winter.

$$\text{Allowance} = \frac{12 \text{ days} \times 3 \text{ winters}}{36 \text{ months} * 22 \text{ days/month}} = 0.047$$

Thus, the overall labor adjustment factor = 1.05 * 1.047 = 1.10

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Source:

* SNC Inc., "Study Relating to the Technical and Economic Feasibility of Lignite Liquefaction in Saskatchewan", October 1982.

Appendix C-3 Machinery, Equipment and Plant Indices

Year	Marshall and Stevens Equipment Cost Index *	Chemical Engineering Plant Indexing *	Chemical and Petrochemical Plant Index **
1974	398.4	165.4	49.7
1975	444.3	182.4	58.8
1976	472.1	192.1	64.0
1977	505.4	204.1	69.0
1978	545.3	218.8	74.4
1979	599.4	238.7	81.8
1980	659.6	261.2	90.2
1981	721.3	297.0	100.0
1982	745.6	314.0	110.2
1983	760.8	316.9	115.5
1984	785.4	344.5	118.3

Sources:

* Chemical Engineering, "Economic Indicators", October 1984.

** Statistics Canada, "Construction Price Statistics", Catalogue

Appendix C-4 Experimental Factors for Capacity Adjustment of Costs of Equipment Items

Equipment Item	Factor
Air Cooler	0.8
Cooling Towers	0.60
Exchangers	
Conventional	0.65
Reboilers	0.50
Filters	0.60 - 0.65
Furnaces	0.80 - 0.85
Generators/Electrical Support	0.75
Motors	0.90 - 1.00
Pressurized Vessels	
Columns and Reactors	0.65 - 0.70
Tanks	0.60 - 0.65
Horizontal Storage Tanks	0.65
Vertical Storage Tanks	0.70
Pump	
Centrifugal	0.40 - 0.50
Reciprocating	0.60 - 0.70
Refrigeration	0.70
Steam Generation	0.80

Source:

Institut Français du Pétrole, "Manual of Economic Analysis of Chemical Process", McGraw Hill Book Company, N.Y., 1981.

Appendix C-5 Typical Manpower Distribution and Average Salary Scale

	% of Total Manpower
Operating Labour (OL)	34
Operation Supervisors (20% of OL)	7
Maintenance Labour (ML)	42
Maintenance Supervision (20% of ML)	8.5
Plant Management and Administrative Support	8.5

Source:
Rogers, K.A., and Hill, R.F., ESCOE, DOE Report No. FE-2468-51, July 1979

APPENDIX - D

ELECTRIC-ARC ACETYLENE REACTOR
(CHEMISCHE WERKE HUELS)

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Appendix D: Electric-Arc Acetylene Reactor (Chemische Werke Huels)

The electric-arc acetylene reactor of Chemische Werke Huels is described in Section 3.1 of the report. Other design features and technical details are given below.

Power Supply: At Huels, the electric-arc is operated with DC primarily due to difficulties associated with normal AC. Either a high frequency AC must be used or the arc must be continuously ignited by special control devices.

The DC is furnished by a grid controlled mercury rectifier. Each reactor has its own adjustable DC supply for ease of regulation. The time constant of the regulating circuit is only a few multiseconds.

For a given current intensity, the arc length and consequently its voltage is determined by regulation of the gas flow. The grid control rectifier adjusts the output voltage so that the predetermined desired values are reached.

In spite of the high voltage (about 7000 volts), no operational difficulties are encountered with the rectifiers, control devices and the electric-arc ignition devices. Operating times of 50 000 hours, sometimes extending to 100 000 hours have been realized.

Electrodes: The life of the electrodes is limited. Cathodes normally are replaced after about 800 hours and anodes after 150 hours. The wall thickness of the electrodes is 10-20 mm. Carbon black produced by the arc deposits on the electrodes form hard graphite which affect the stability of the arc.

Appendix D: Electric-Arc Acetylene Reactor (Chemische Werke Huels) (Cont'd)

Gas Flow and Temperature: The temperature of the arc itself is in the range of 18 000 to 20 000 C. On the other hand, the temperature of the inner wall of the water-cooled electrode is about 500 C, therefore a steep radial temperature gradient exists in the arc zone.

Rapid heat transfer and high energy transmission per cm of arc length is achieved mainly due to:

- high temperature of the electric-arc itself
- high diffusion velocities and thermal conductivity of the hydrogen atoms and ions formed which can also rapidly transfer association, ionization and excitation energy to other reading species.